U.S. DEPARTMENT OF COMMERCE
National Technical Information Service
PB80-108970

Chemical Primary Sludge Thickening and Dewatering

Envirotech Corp, Salt Lake City, UT Eimco Process Macinery Div

Prepared for

Municipal Environmental Research Lab, Cincinnati, OH

Jul 79

United States
Environmental Protection
Agency

Municipal Environmental Research
Laboratory
Cincinnati OH 45268

EPA-600/2-79-055
July 1979

Research and Development

EPA

Chemical Primary Sludge Thickening and Dewatering

PB80-108970

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TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO.
EPA-600/2-79-055

2.

3. RECIPIENT'S ACCESSION NO.
PB 80 108970

4. TITLE AND SUBTITLE
CHEMICAL PRIMARY SLUDGE THICKENING
AND DEWATERING

5. REPORT DATE
July 1979 (Issuing Date)

6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S)
David DiGregorio, J. Brian Ainsworth, and
Keith J. Mounteer

8. PERFORMING ORGANIZATION REPORT NO.

9. PERFORMING ORGANIZATION NAME AND ADDRESS
Envirotech Corporation
Eimco Division
Salt Lake City, Utah 84110

10. PROGRAM ELEMENT NO.
1BC611 SOS 1 Task A/18

11. CONTRACT/GRANT NO.
68-03-0404

12. SPONSORING AGENCY NAME AND ADDRESS
Municipal Environmental Research Laboratory--Cin.,OH
Office of Research and Development
U. S. Environmental Protection Agency
Cincinnati, Ohio 45268

13. TYPE OF REPORT AND PERIOD COVERED

14. SPONSORING AGENCY CODE
EPA/600/14

15. SUPPLEMENTARY NOTES
Project Officer: Roland Villiers (513)684-7664

16. ABSTRACT

This report presents the results of a ten month study of the thickening and dewatering characteristics of chemical-primary sludges. Alum-primary and ferric-primary sludges were produced in parallel trains of a pilot plant operated using a municipal wastewater. Each chemical treatment unit was operated under several coagulant doses during the four phases of this study resulting in the production of several chemical-primary sludges with distinct characteristics.

Gravity thickening and dissolved air flotation thickening results for each chemical-primary sludge are presented. Gravity thickening was evaluated using continuous, pilot scale gravity thickeners; dissolved air flotation thickening evaluations were performed using batch, bench-scale equipment. Sludge dewatering evaluations were performed for all chemical-primary sludges using a pilot scale bowl centrifuge, vacuum belt filter and filter press.

The report presents correlations developed relating performance of each unit operation to specific characteristics identified for each chemical-primary sludge. An economic analysis of centrifugation and vacuum belt filtration of each chemical-primary sludge is presented.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Phosphorus, Sludge, Sludge disposal, Sewage treatment, Centrifuges, Dewatering, Thickening, Filter press, Economic analysis, Phosphate, Alums	Phosphorus sludges, Phosphate removal, Wastewater treatment, Ferric chloride, Air flotation, Vacuum filter,	13B

18. DISTRIBUTION STATEMENT Release to public	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 290
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE A13-A01

EPA Form 2220-1 (9-73)

☆ U.S. GOVERNMENT PRINTING OFFICE: 1979-657-060/5429

EPA-600/2-79-055
July 1979

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

This report summarizes the results of a pilot plant program which developed application and design criteria for thickening and dewatering waste sludges produced from chemical clarification of municipal wastewater. The development of such information provides valuable insight to engineers in their efforts to design efficient, cost-effective wastewater treatment systems.

Francis T. Mayo
Director
Municipal Environmental
Research Laboratory

ABSTRACT

This report presents the results of a ten month study of the thickening and dewatering characteristics of chemical-primary sludges. Alum-primary and ferric-primary sludges were produced in parallel trains of a pilot plant operated using a municipal wastewater. Each chemical treatment unit was operated under several coagulant doses during the four phases of this study resulting in the production of several chemical-primary sludges with distinct characteristics.

Gravity thickening and dissolved air flotation thickening results for each chemical-primary sludge are presented. Gravity thickening was evaluated using continuous, pilot scale gravity thickeners; dissolved air flotation thickening evaluations were performed using batch, bench-scale equipment. Sludge dewatering evaluations were performed for all chemical-primary sludges using a pilot scale solid bowl centrifuge, vacuum belt filter and filter press.

The report presents correlations developed relating performance of each unit operation to specific characteristics identified for each chemical-primary sludge. An economic analysis of centrifugation and vacuum belt filtration of each chemical-primary sludge is presented.

This report was submitted in fulfillment of EPA Contract Number 68-03-0404, by the Eimco Process Machinery Division of Envirotech Sanitary Engineering Technology Department, under the (partial) sponsorship of the Environmental Protection Agency. Work was completed as of January 1976.

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ABBREVIATIONS AND SYMBOLS

Al_r	aluminum available after its utilization in $AlPO_4$, mg/ℓ
alum	aluminum sulfate
A/S	air to solids ratio, weight percent
BOD_5	5-day biochemical oxygen demand, mg/ℓ
C	conditioned feed concentration to filter press, g/ℓ
^{o}C	degrees Celsius
Ci	thickener feed concentration, g/ℓ
Cu	thickener underflow concentration, g/ℓ
COD	chemical oxygen demand, mg/ℓ
CT	chemical treatment unit - an abbreviation for flocculating clarifier
CT-A	flocculating clarifier in which alum was the coagulant
CT-F	flocculating clarifier in which ferric chloride was the coagulant
D_1	cake dry solids weight, g
DAF	dissolved air flotation
DC	direct current
DS	dry suspended solids
E	excellent discharge of vacuum filter cake
F	fair discharge of vacuum filter cake

ABBREVIATIONS AND SYMBOLS (continued)

$^{\circ}F$	degrees Fahrenheit
Fe_r	iron available after its utilization in $FePO_4$, mg/ℓ
FFR	form filtration rate, Kg/hr-sq m
FSFR	full-scale filtration rate, Kg/hr-sq m
Ferric Chloride	a source of Fe^{+++} in liquid form
G	centrifugal force when used in centrifuge description
G	good discharge of vacuum filter cake when used in a vacuum filter description
HLR	hydraulic loading rate, cu m/sq m-hr
JTU	Jackson Turbidity Units
kw	kilowatt
ND	non-dischargeable vacuum filter cake
O&M	operating and maintenance
P	poor discharge of vacuum filter cake
Phase	a time period under which flocculating clarifier was operated at selected chemical and/or polymer dosages
pH	negative logarithm of the hydrogen ion concentration
Q	liquid flowrate
ΔP	difference in feed and filtrate densities, g/ℓ
rpm	revolutions per minute
Δrpm	difference between centrifuge bowl rpm and conveyor rpm
S_c	cake dry solids, weight percent
$SBOD_5$	soluble 5-day biochemical oxygen demand, mg/ℓ

ABBREVIATIONS AND SYMBOLS (continued)

SCOD	soluble chemical oxygen demand, mg/ℓ
SLR	solids loading rate, Kg/sq m-hr
TDS	ton of dry suspended solids
TS	total solids, g/ℓ
TSL	thickener solids loading, Kg/day-sq m
TSS	total suspended solids, mg/ℓ
θ	a measurement of time
θ_D	vacuum filter dry time, minutes
$\theta_{D/W}$	a correlating factor used in describing vacuum filter performance, min-sq m/Kg
θ_F	vacuum filter form time, minutes
ΔV	final filtrate volume minus filtrate volume at cycle time, t
W	cake dry weight per unit area, Kg/sq m
W_1	wet cake weight at end of test, g

ACKNOWLEDGEMENTS

The support and indulgence of the Salt Lake City Commissioner of Water and Sewers, city engineers and municipal sewage pumping plant personnel is acknowledged.

Special acknowledgement is due Messrs. Doug Hendry, Steve Nelson, Jack Otto, and Charles Thompson who nursed the pilot plant through start-up and kept it operating. Mr. Hendry was of great assistance in preparing the figures.

The support, counsel and patience of the Project Officer, Mr. Roland Villiers, and Dr. Joseph Farrell of the Water Research Division, Municipal Environmental Research Laboratory, Environmental Protection Agency, Cincinnati, Ohio, is acknowledged with sincere thanks.

SECTION I

INTRODUCTION

The discharge of phosphorus-containing wastewater into the surface water of the United States has contributed to their overfertilization and eutrophication. As a result, efforts are now being made to remove phosphorus from wastewater.

Phosphorus removal from raw wastewater is normally achieved by precipitating it from solution with a metal salt of iron, aluminum or calcium. The use of phosphorus removal metal not only increases the amount of sludge produced but the character of sludge is changed. This results in thickening and dewatering properties different from raw-primary sludge. It is important that means of thickening and dewatering these combined chemical-primary sludges be investigated and determined so equipment sizing parameters can be determined.

In March, 1974, EPA and the Eimco Process Machinery Division of Envirotech Corporation initiated a research contract to operate a pilot plant to generate various chemical-primary sludges, and to obtain performance data describing the thickening and dewatering characteristics of these sludges. The project, entitled "Chemical-Primary Sludge Thickening and Dewatering," was conducted in facilities originally constructed in 1969 where two prior major studies on physical-chemical treatment of municipal wastewater had been conducted.[1,2] This latest study was conducted by generating chemical-primary sludges, produced from either alum addition or ferric chloride addition to effect phosphorus removal from municipal wastewater, thickening the resultant underflow sludge in gravity thickeners, and then dewatering it by vacuum filtration or centrifugation (and in a few instances by pressure filtration).

The general objectives of the study, as originally established, were as follows:

1. To determine the thickening and dewatering characteristics of chemical-primary sludges generated from chemical clarification of municipal wastewater using alum or ferric chloride, each at two levels of phosphorus removal.

2. To establish sludge conditioning chemical requirements for dewatering alum and ferric-primary sludges by vacuum filtration and centrifugation.

3. To perform a comparison of centrifugal and vacuum filtration dewatering of chemical-primary sludges generated from each operating condition during this study.

An important product of the investigation was to be an evaluation of the impact of dewatering cost on the phosphate removal by Fe^{+++} or Al^{+++} salts.

SECTION II

CONCLUSIONS

1. Sludge production in the chemical-primary flocculating clarifier was predicted with reasonable accuracy from analysis of physical and chemical constituents around the chemical-primary unit and knowledge of chemical coagulant dose.

2. No evidence was found to indicate that the addition of polymer to the primary chemical treatment unit had any effect on subsequent sludge thickening or dewatering operations.

3. Dissolved air flotation thickening produced satisfactory sludge solids concentration and solids capture without the need for chemical flotation aids.

4. Dissolved air flotation thickening performance was adversely affected by the presence of aluminum or iron chemical solids in the feed sludge to the flotator. Allowable loading rates and float solids concentration generally decreased as the amount of chemical solids in the feed sludge increased.

5. Dissolved air flotation thickening performance characteristics of alum-primary sludge were superior to those of ferric-primary sludge at equivalent levels of phosphorus removal of 80 and 95 percent.

6. Dissolved air flotation thickening of alum-primary sludge, ferric-primary sludge and primary sludge produced thickened sludge concentrations of 2.8-5.5, 3.5-4.5 and 7.0-8.0 weight percent, respectively depending upon the amount of chemical solids present in the flotator feed sludge, and the air to solids ratio chosen.

7. Gravity thickener underflow solids concentration was adversely affected by the presence of aluminum or iron chemical solids in the feed sludge to the thickener. Underflow solids concentration generally decreased as the amount of chemical solids in the feed sludge increased. The decrease in thickener underflow solids concentration with increasing chemical solids in the feed sludge was more significant for alum-primary sludge than ferric-primary sludge.

8. Gravity thickener performance characteristics of ferric-primary sludge were superior to those of alum-primary sludge at equivalent levels of phosphorus removal of 80 and 95 percent.

9. Thickened alum-primary sludge solids concentrations from flotation thickening were higher than those achieved with gravity thickening while the reverse was true for ferric-primary sludge.

10. Gravity thickening of alum-primary, ferric-primary and primary sludge produced thickened sludge concentrations of 2.5-4.5, 4.5-6.0 and 7.0-10.0 weight percent, respectively, depending upon the amount of chemical solids present in the thickener feed sludge and the thickener operating conditions chosen.

11. Vacuum filtration performance, relative to required chemical conditioning dose, filtration rate and filter cake solids content, was adversely affected by the presence of aluminum chemical solids in the feed sludge to the filter. The required chemical conditioning dose increased and the filtration rate decreased as the quantity of aluminum chemical solids in the feed sludge increased. Cake solids content was insensitive to the amount of aluminum chemical solids present in the filter feed sludge with the exception that primary sludge dewatered to cake solids content levels higher than alum-primary sludge.

12. Vacuum filtration performance, relative to required chemical conditioning dose and filtration rate, was adversely affected by the presence of iron chemical solids in the feed sludge to the filter. In general, required chemical conditioning dose increased and filtration rate decreased as the quantity of iron chemical solids in the feed sludge increased. Cake solids content was insensitive to the amount of iron chemical solids present in the filter feed sludge with the exception that ferric-primary sludge dewatered to cake solids content levels higher than primary sludge.

13. Volatile solids, phosphorus and the primary coagulant metal were efficiently captured during vacuum filtration of alum-primary or ferric-primary sludge.

14. Vacuum filtration of alum-primary, ferric-primary and primary sludge produced filter cakes of 25-27, 34-35 and 26-29 weight percent total solids, respectively, depending upon the filter operating conditions chosen.

15. Centrifugal dewatering performance, relative to polymer requirements and cake solids concentrations, was adversely affected as the quantity of aluminum chemical solids present in the feed sludge increased. Machine capacity, to achieve a given level of solids capture, was not significantly affected by the presence of aluminum chemical solids in the feed sludge.

16. Centrifugal dewatering performance, relative to polymer requirements and machine capacity to achieve a given level of solids capture, was not affected by the quantity of iron chemical solids present in the feed sludge. Cake solids concentrations increased as the amount of iron chemical solids present in the feed sludge increased.

17. Centrifugal dewatering of ferric-primary sludge produced cake solids concentrations significantly higher than those achieved with alum-primary sludge.

18. Centrifugal dewatering of alum-primary, ferric-primary and primary sludge produced cakes of 15-18, 22-25 and 20-21 weight percent total solids, respectively, at total solids capture of approximately 95 percent.

19. Volatile solids, phosphorus and the primary coagulant metal were efficiently captured during centrifugal dewatering of alum-primary or ferric-primary sludge at 90-95 percent total solids capture.

20. Pressure filtration cake solids concentrations were adversely affected by the amount of aluminum chemical solids in the feed sludge to the press.

21. Presure filtration of alum-primary sludge produced cakes ranging from 25-35 weight percent total solids and 30-41 weight percent total solids for primary sludge depending upon cake thickness, cycle time, and the amount of aluminum chemical solids present in the feed sludge to the press.

22. It was costlier by 10-20 percent to centrifugally dewater and 15-25 percent to vacuum filter dewater alum-primary or ferric-primary sludge produced from chemical treatment aimed at 95 percent phosphorus removal as compared to 80 percent phosphorus removal. This was caused by the increased quantities of sludge and a generally more difficult sludge to dewater for the 95 percent phosphorus removal case.

23. There was no significant difference between centrifugal and vacuum filtration dewatering costs for either alum-primary or ferric-primary sludge produced from chemical treatment aimed at 95 percent phosphorus removal. The

total costs (dewatering plus cake incineration) were 25-30 percent higher when centrifugation was used as compared to vacuum filtration.

24. Centrifugal dewatering was 10-15 percent more expensive than vacuum filtration of either alum-primary or ferric-primary sludge produced from chemical treatment aimed at 80 percent phosphorus removal. The total costs (dewatering plus cake incineration) were 10-35 percent higher when centrifugation was used as compared to vacuum filtration for the alum-primary sludge case. The total costs (dewatering plus cake incineration were 30-45 percent higher when centrifugation was used as compared to vacuum filtration for the ferric-primary sludge case.

25. The total costs for disposal (dewatering plus cake incineration) of alum-primary sludge produced from 95 percent phosphorus removal were approximately 25 percent higher when centrifugation was used as compared to vacuum filtration. The cost differential increased to approximately 30 percent for the case of ferric-primary sludge produced from 95 percent phosphorus removal.

26. The total costs for disposal (dewatering plus cake incineration) of alum-primary sludge produced from 80 percent phosphorus removal averaged approximately 20 percent higher when centrifugation was used as compared to vacuum filtration. The cost differential increased to approximately 40 percent for the case of ferric-primary sludge produced from 80 percent phosphorus removal.

SECTION III

RECOMMENDATIONS

1. Operating control techniques should be developed for preventing addition of excessive amounts of primary coagulant (alum or ferric chloride) to raw wastewater for specific levels of phosphorus removal. The need for these controls was demonstrated during this study from documented deterioration in thickening and dewatering properties of chemical-primary sludges as the dosage of primary coagulant increased.

2. Further studies would appear to be warranted in the area of gravity thickening of chemical-primary sludges using organic flocculants.

3. Further study of alternative dewatering devices should be performed, particularly pressure filtration and horizontal belt filtration.

SECTION IV

EXPERIMENTAL SYSTEM

PILOT PLANT EQUIPMENT

The pilot facility consisted of two identical chemical treatment and sludge thickening trains. Raw wastewater, following degritting in a hydrocyclone and screening through a 12.7 mm (.50 inch) opening mesh screen, was contacted with alum in one train and with ferric chloride in the other. The chemical primary sludges produced were gravity thickened. The sludges were then dewatered by vacuum filtration, centrifugation, or pressure filtration. A schematic of the pilot plant flowsheet is shown on Figure 1.

An auxiliary line tapped into a discharge header of the Metropolitan Salt Lake City pump station supplied raw wastewater to the pilot plant. Approximately 1000 ℓ/minute (265 gpm) of raw wastewater was pumped through a 15.2 cm (6 inch) diameter Hydrocyclone* for grit removal. The degritted wastewater was discharged to a 1500 ℓ (400 gallon) wet well and screened through a 12.7 mm (0.50 inch) opening wire mesh screen. Since only 150-300 ℓ/minute (40-80 gpm) of degritted screened wastewater was used for pilot plant operations, the majority of the flow entering the wet well overflowed to a drain and returned to the pump station. Hydrocyclone operation was continuous, whereas routine, manual cleaning of the wire mesh was required.

Degritted, screened wastewater was pumped from the wet well to the pilot plant where the flow was split into two streams. Control of the flow to each chemical treatment unit was achieved by adjustments of the feed pump speed and was fine-tuned by throttling gate valves installed in each line.

Feed flow measurement for the ferric chemical treatment unit (CT-F) was achieved using a magnetic flow meter. The magnetic flow meter signal was directed to a recording flow rate indicator and flow totalizer.

Because of continual delays in delivery of a second magnetic flow meter, an orifice plate and mercury manometer were used to

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*Wemco Division, Envirotech Corporation

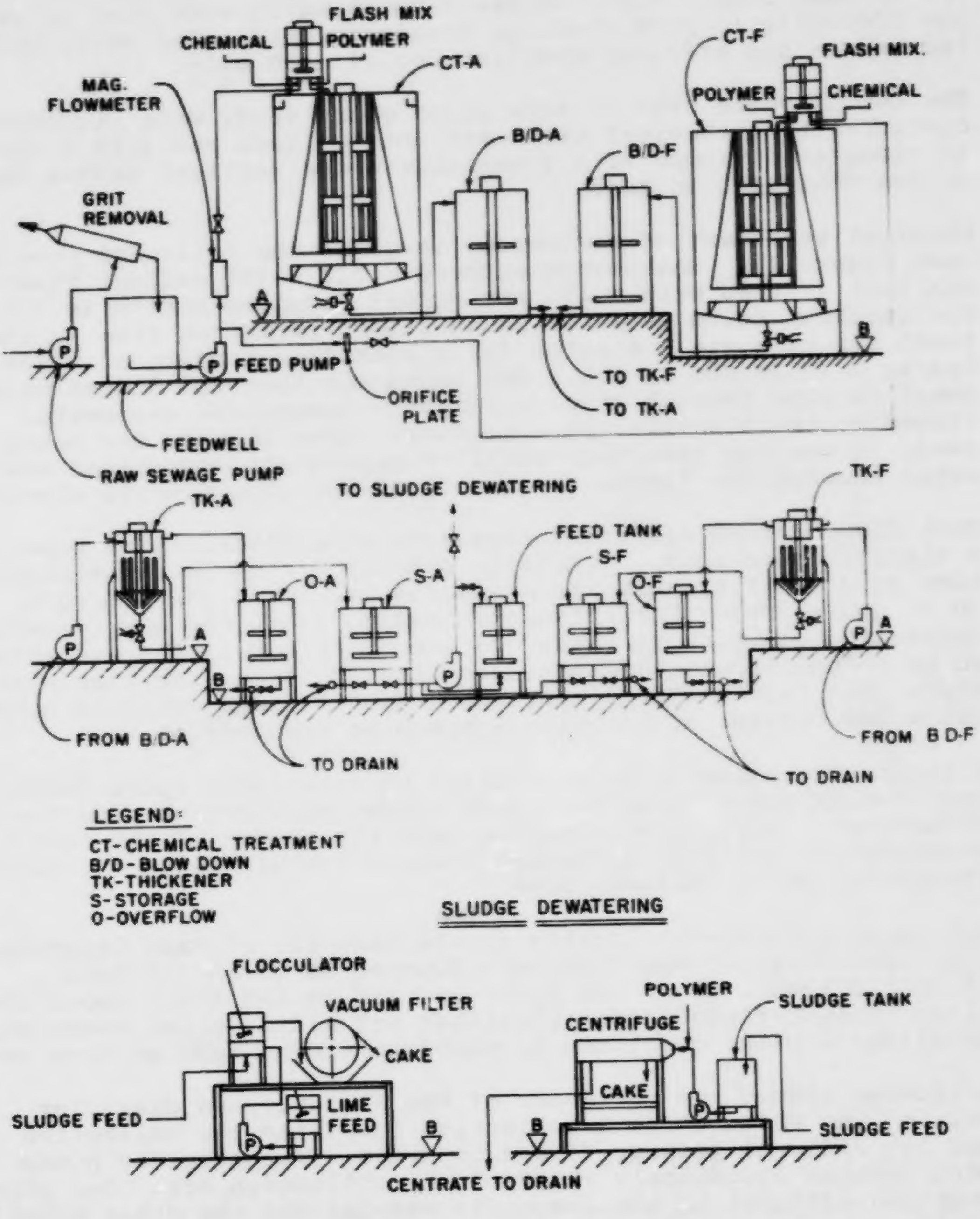
FLASH MIX
CHEMICAL
POLYMER
CT-A
MAG. FLOWMETER
GRIT REMOVAL
B/D-A
B/D-F
CT-F
FLASH MIX.
POLYMER
CHEMICAL
P
P
FEED PUMP
ORIFICE PLATE
A
B
TO TK-F
TO TK-A
FEEDWELL
RAW SEWAGE PUMP
TK-A
TO SLUDGE DEWATERING
FEED TANK
TK-F
O-A
S-A
S-F
O-F
P
A
B
FROM B/D-A
TO DRAIN
TO DRAIN
FROM B/D-F
LEGEND:
CT-CHEMICAL TREATMENT
B/D-BLOW DOWN
TK-THICKENER
S-STORAGE
O-OVERFLOW
SLUDGE DEWATERING
FLOCCULATOR
VACUUM FILTER
CAKE
SLUDGE FEED
LIME FEED
P
B
POLYMER
SLUDGE TANK
CENTRIFUGE
CAKE
B
P
SLUDGE FEED
CENTRATE TO DRAIN

FIGURE 1. PILOT PLANT FLOWSHEET SCHEMATIC.

measure the rate of flow in the alum chemical treatment unit (CT-A) feed line. Sixty degree V-notch weirs were used to verify the flow rate to each chemical treatment unit. The weirs were located in the effluent overflow box of each unit.

The two CT units used in this pilot plant study were flocculator-clarifier type chemical treatment units. Each was 3.05 m (10 ft) in diameter equipped with a variable-speed vertical paddle wheel in the flocculation zone.

Chemical treatment of wastewater involved the following processes (see Figure 2). Wastewater entered a 210 ℓ (55 gallon) flash mix tank (fitted with a 175 rpm mixer) from the bottom as did the chemical being used for coagulation; retention time in the flash mix tank was 2 minutes for a clarifier loading of 0.204 ℓps/sq m (0.20 gpm/sq ft). The rapid mix tank was fitted with an overflow pipe through which chemically coagulated wastewater flowed to the flocculating clarifier. When polymer was being used, it was fed into this overflow pipe where coagulated wastewater entered the flocculation zone of the flocculating clarifier.

Each flocculating clarifier consisted of a flocculation zone and a clarification zone. Hydraulic retention time in the flocculation zone was 1.8 hours for a loading rate of 0.204 ℓpm/sq m (0.30 gpm/sq ft). For this same loading rate, the clarification hydraulic retention time was 4.8 hours. Flocculation was effected by a "paddle wheel" mounted vertically inside the flocculation zone. The paddle flocculator was connected to a variable speed drive for control of rotational speed as required.

A raking mechanism (also controlled by a variable speed drive) continually brought the thickened sludge to a center cone for withdrawal. A plate was mounted directly on top of the raking mechanism to minimize hydraylic-induced turbulence in the sludge thickening and withdrawal zone.

Four sample taps were located in the sidewall of each flocculating clarifier. These taps were located at 36 cm (14 inch), 79 cm (31 inch), 99 cm (39 inch) and 150 cm (59 inch) above the floor of each flocculating clarifier and aided in its operation by allowing plant operators to withdraw sludge samples from them.

A launder around the perimeter of the flocculating clarifier carried the effluent to a collection box; from the collection box the effluent gravity flowed to drain. Two sampling pipes were located immediately ahead of the collection box. One pipe supplied effluent to the composite sampler and the other supplied effluent to the turbidimeter. On the discharge side of the collection box, a V-notch weir was located. Whenever weir measurements were taken to check the flow rate, both of the sample pipes were closed.

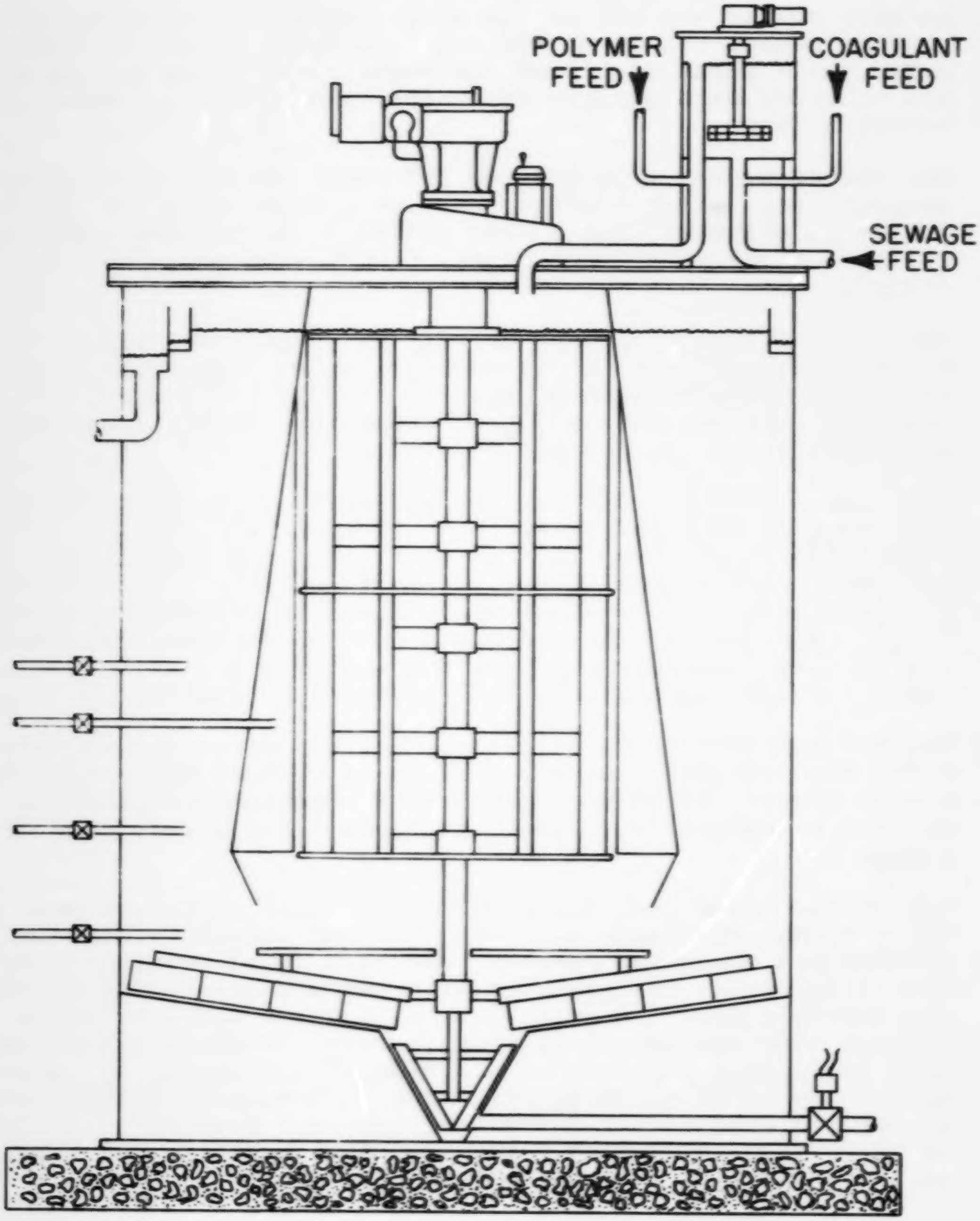
POLYMER FEED
COAGULANT FEED
SEWAGE FEED

FIGURE 2. FLOCCULATING CLARIFIER TREATMENT UNIT.

Thickened sludge was removed from each CT unit and fed to a gravity thickener on a timed basis. The thickener for ferric-primary sludge was 107 cm (42 inch) diameter, and that for alum primary sludge was 91 cm (36 inch) diameter. Each thickener had a side wall depth of 152 cm (60 inch), a 45° cone on the bottom, was equipped with pickets and rakes, and had an automatic desludging valve.

The overflow from each gravity thickener flowed to a separate storage tank. Each storage tank was 152 cm (60 inch) diameter with a capacity of 3560 ℓ (940 gallon). As required, the contents of these tanks were mixed, inventoried and manually drained.

Thickened sludge from each gravity thickener was removed on a timed basis and retained in storage tanks. Each tank was 152 cm (60 inch) diameter with a capacity of 2910 ℓ (770 gallon). As required, the contents of these tanks were mixed, inventoried and transferred for dewatering studies.

The vacuum filter* used in the dewatering studies was 91.4 cm (36 inch) in diameter by 30.5 cm (12 inch) wide with a continuous belt. The filter drum and vat were made of a thermoplastic material. The vacuum filter, with flocculator, vacuum pump, filtrate pump, and receiver was mounted on a common platform. A 210 ℓ (55 gallon) tank equipped with a vertical rapid mixer and positive displacement pump coupled to a variable speed drive served as the conditioning lime slurry feed system.

Centrifugal dewatering of all sludges generated during this study was evaluated using a 15.2 cm (6.0 inch) diameter, horizontal scroll, solid bowl machine.** Pertinent centrifuge machine parameters are listed in Table 1 and illustrated on Figure 3.

The filter press used was a 30 cm (12 inch) polypropylene, center feed, corner discharge unit with ratchet closing. Maximum recommended operating pressure was 10.5 Kg/sq cm (150 lb/sq in). The filter press was equipped with two chambers; cake thickness was variable from 2.5 cm to 3.8 cm (1.0-1.5 inch) by using various combinations of chamber spacers. Sewn-center cloths were used, with the area of one side of a cloth being approximately 500 sq cm (0.54 sq ft). The skid-mounted filter press system included an air compressor (maximum pressure = 14 Kg/sq cm (200 lb/sq in)) which functioned to pressurize the 115 ℓ (30 gallon) feed tank displacing feed sludge into the press chambers.

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*Eimco-Belt Filter, Eimco PMD of Envirotech Corporation, Salt Lake City, Utah

**P600 Sharples Super Decanter, Sharples Division of Pennwalt Corporation, Warminster, Pennyslvania

TABLE 1

CENTRIFUGE MACHINE PARAMETERS

Sharples Model P600

Parameter	Values
Bowl diameter (cm)	15.2
Pool length* (cm)	31.4
Pool depth* (cm)	1.27
Beach length* (cm)	2.54
Beach angle (degrees)	10.0
Pool volume* (ℓ)	1.56
Conveyer pitch (cm)	5.08
Bowl rpm	5000
Nominal G	2100
Differential rpm	10-25

*Corresponds to No. 3 pool setting

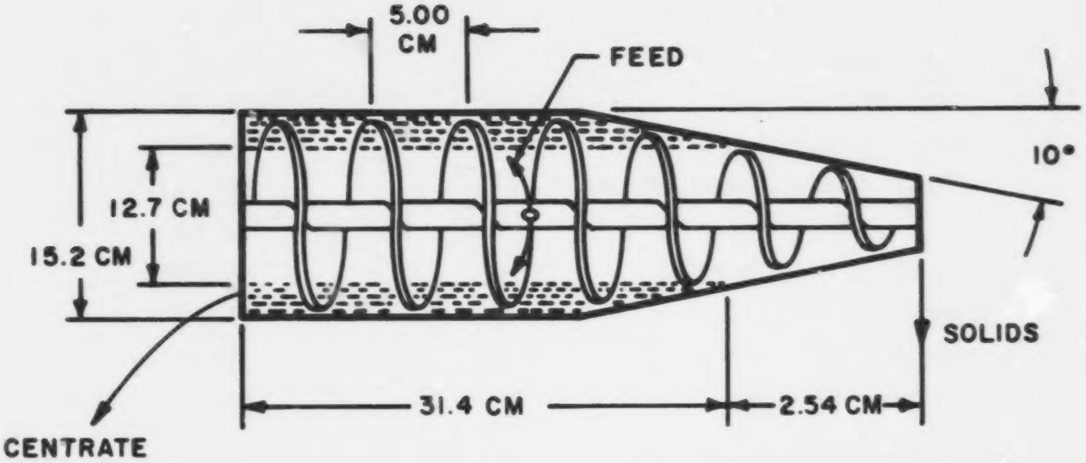
5.00 CM
FEED
10°
12.7 CM
15.2 CM
SOLIDS
31.4 CM
2.54 CM
CENTRATE

FIGURE 3. SCHEMATIC OF SOLID BOWL CENTRIFUGE.

The feed tank was mounted on a "rocking" mechanism which was used to keep the sludge solids in suspension. An air pressure regulator was located in the line between the compressor and feed tank to provide for a controlled pressure build-up as desired.

BENCH-SCALE EQUIPMENT

Dissolved air flotation (DAF) thickening was evaluated using a batch, bench-scale dissolved air flotation device. DAF thickening evaluations were performed on all sludges generated during this pilot plant study. A 61.5 sq cm (0.066 sq ft) bench-scale batch DAF device, as shown on Figure 4, was used for the evaluations. The flotation device consisted of a calibrated 1 ℓ flotation chamber, pressurization chamber and ancillary pressure controls, piping and sampling tap.

Vacuum filtration of the primary sludge generated during Phase 4 was evaluated using a bench-scale, 93 sq cm (0.1 sq ft) filter leaf. As schematically shown on Figure 5, the test apparatus consisted of a vacuum source and receiver, filter leaf and sludge slurry container.

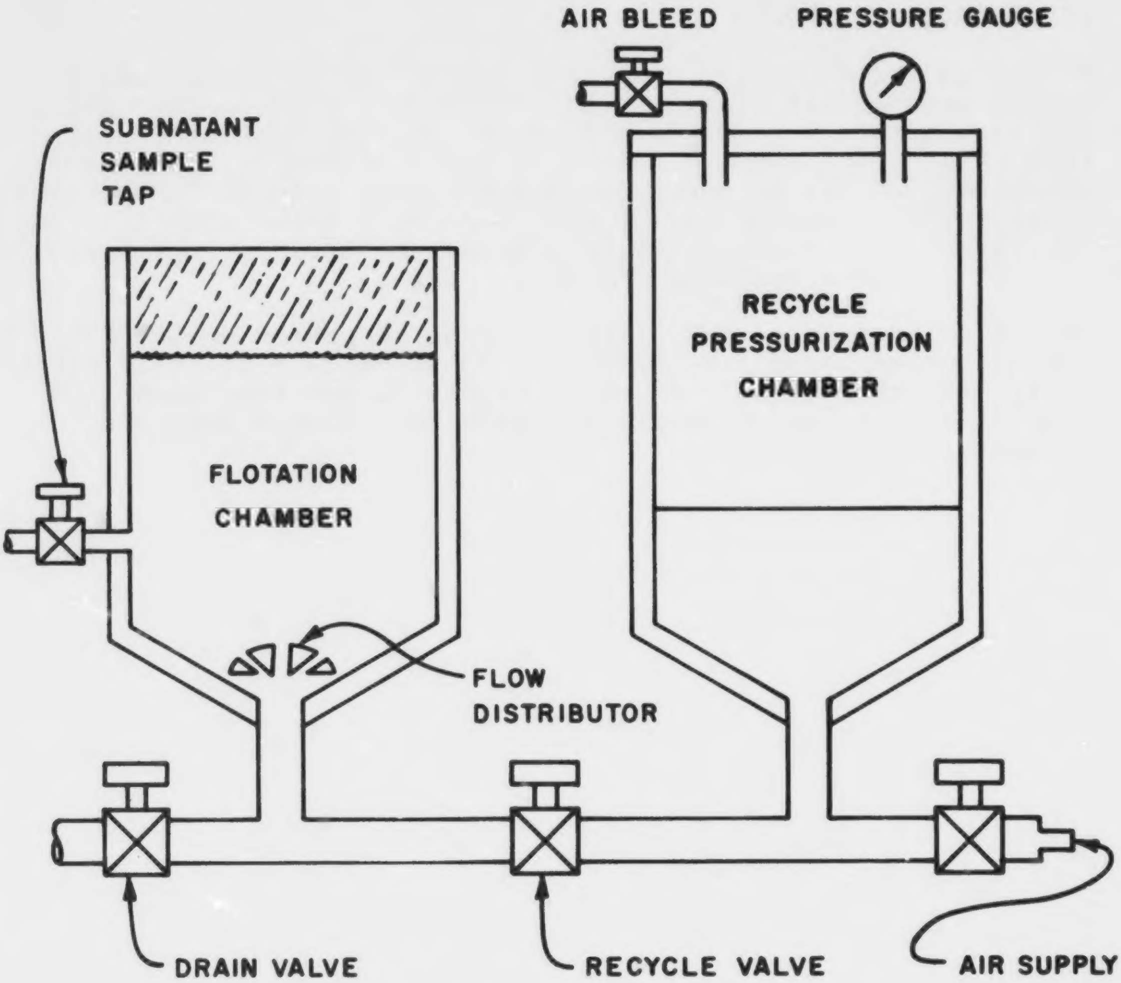
AIR BLEED
PRESSURE GAUGE
SUBNATANT SAMPLE TAP
RECYCLE PRESSURIZATION CHAMBER
FLOTATION CHAMBER
FLOW DISTRIBUTOR
DRAIN VALVE
RECYCLE VALVE
AIR SUPPLY

FIGURE 4. DISSOLVED AIR FLOTATION THICKENING TEST APPARATUS.

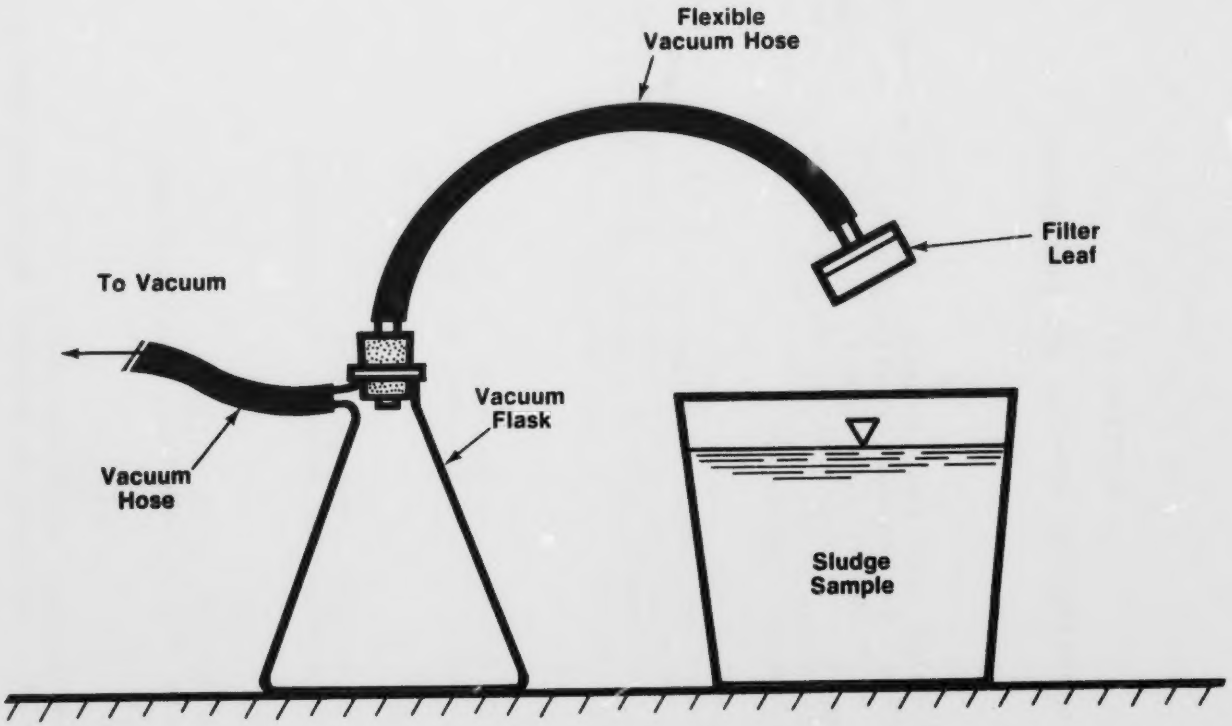
Flexible
Vacuum Hose
Filter
Leaf
To Vacuum
Vacuum
Flask
Vacuum
Hose
Sludge
Sample

FIGURE 5. BENCH SCALE FILTER LEAF TEST APPARATUS.

SECTION V

EXPERIMENTAL METHODS AND PROCEDURES

GENERAL

The study was divided into four phases. Each phase consisted of chemically treating raw wastewater with alum in one train of the pilot plant and with ferric chloride in the other. Variations in the properties of the chemical-primary sludges produced were achieved by varying the chemical dosage and/or with the use of polymer during chemical treatment.

The approach used in pilot plant operation under each phase involved establishment of stable chemical treatment operating conditions followed by pilot plant equipment performance evaluations. Chemical treatment operating conditions were established in accordance with the following specific program objectives:

Phase 1 (P-1)

1. High alum dosage, no polymer
2. High ferric chloride dosage, no polymer

Phase 2 (P-2)

1. High alum dosage, with polymer
2. High ferric chloride dosage, with polymer

Phase 3 (P-3)

1. Low alum dosage, with polymer
2. Low ferric chloride dosage, with polymer

Phase 4 P-4)

1. High alum dosage, with polymer
2. No chemical addition, no polymer addition; primary sludge

High chemical dosage was defined as the minimum chemical dosage adequate to achieve approximately 95 percent phosphorus removal. Low chemical dosage was defined as the minimum chemical dosage adequate to achieve approximately 80 percent phosphorus removal. However, experience has shown that a chemical dosage sufficient to effect 80 percent phosphorus removal could result in poor effluent clarity. Therefore, either 80 percent phosphorus removal or acceptable effluent clarity was established as the criterion for determining the low chemical dosage. From previous studies[2], an effluent suspended solids concentration of 30 mg/ℓ or less was considered acceptable.

PILOT PLANT PROCESS CONTROL PROCEDURES

Chemical Treatment Unit Operation

Wastewater was metered to a flash mix tank ahead of each chemical treatment unit where it was contacted with inorganic coagulant. Either liquid alum or liquid ferric chloride was used. The liquid alum (8.25 weight percent as Al_2O_3) was stored in a 2500 ℓ (660 gallon), rubber lined, steel storage tank, and fed to the alum chemical treatment unit (CT-A) using a diaphragm pump* driven by a variable speed DC motor. The liquid ferric chloride (15.2 weight percent as Fe) was stored and fed in an identical manner.

When polymer was used, it was added immediately before the contents of the flash mix tank entered the flocculation zone of the chemical treatment unit. Polymer was prepared by mixing approximately 200 g of anionic polymer with 200 ℓ (53 gallon) water in a 210 ℓ (55 gallon) container and vigorously agitating for one-half hour. Following mixing, polymer solution was gravity drained to a 210 ℓ (55 gallon) container where it was metered to the chemical treatment unit using a pump identical to that used for inorganic chemical feeding.

Feed was pumped from the wet well to each flocculating-clarifier by a centrifugal pump controlled by a DC motor. A magnetic flowmeter was installed in one feed line, and an orifice plate with mercury manometer was installed in the other. Sixty degree V-notch weirs were located in the discharge box of each flocculating-clarifier.

By setting the proper flow through the feed line equipped with the magnetic flowmeter, effluent flow over the V-notch weir was noted and this measurement was used to set the same flow to the unit without a magnetic flowmeter installed in the feed line. When installation of a second magnetic flowmeter proved impractical, due to a long delivery time, an orifice plate and mercury manometer were installed in the feed line. This system worked well; sewage flow was checked several times per day by viewing the manometer. The V-notch weirs were then used as a secondary check.

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*Wallace and Tiernan, Model 44-213

Flocculation within the CT unit was developed by slow rotation of the paddle assembly in the baffled flocculation chamber. The flocculated wastewater flowed out from the bottom of the flocculation chamber into a clarification chamber. A portion of the overflow from each CT unit flowed through separate turbidimeters* which continuously measured and recorded effluent turbidity.

Chemical treatment unit operation consisted mainly of routine monitoring of feed, chemical and polymer flow rates. Attempts were made to maintain a constant sludge blanket level in each CT unit. Samples were routinely withdrawn from the sample taps located in the sidewall of the clarification zone and settled in 2 ℓ graduated cylinders. Sludge volume after a 10 minute settling period was recorded for each sample. The results of this test were used by the operators to adjust the timer-controlled sludge blowdown system. As the operators gained experience in interpreting these settling tests, they were able to predict (within one-half percent) the underflow solids concentration removed from the flocculating clarifier. The basic objective in attempting to maintain a constant sludge blanket level in each CT unit was to retain the sludge in the thickening zone for as short a period of time as practical, consistent with achieving reasonably thick underflow sludge. In practice, maintaining a sludge blanket depth of approximately one foot at the sidewall resulted in a solids retention time of less than one day without measureable diminution of underflow concentration due to the minimal sludge depth.

Thickener Operation

Each chemical treatment unit was equipped with a timer-controlled, pneumatic actuated de-sludge valve that allowed sludge to be withdrawn and flow to a surge tank. From the surge tank, sludge was fed directly to the thickener, again on a timed basis; the contents of each surge tank were mixed at timed intervals coincident with sludge drainoff from each surge tank.

Pilot plant operation during most of the phases was such that more sludge was produced than could be thickened with the existing equipment. Therefore, when either surge tank became full, its contents were inventoried (for subsequent sludge production calculations) and wasted.

Sludge was pumped to each gravity thickener on an intermittent, timed basis. The thickener feeding schedule was modified as required to maintain a predetermined solids loading rate to each thickener by selecting the frequency and duration of the feed schedule with electric timers. The operating procedure established for gravity thickener operation involved routine sludge blanket level monitoring with thickener sludge blowdown timer adjustments as necessary to maintain a constant thickener sludge blanket level.

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*Hach, Model 1720 Low Range

Both thickener underflow and supernatant flows were inventoried in separate storage tanks to permit calculation of pertinent thickener operating and performance parameters.

SLUDGE THICKENING AND DEWATERING PROCEDURES

Dissolved Air Flotation Thickening

Bench-scale dissolved air flotation (DAF) thickening evaluations were performed on all sludges generated during this pilot plant study. Chemical flotation aids were not used at any time during the DAF study.

Process supernatant was aerated at 4.6 Kg/sq cm (65 lb/sq in) gauge pressure for approximately 15 minutes within the pressurization chamber in order to saturate the liquid with air. A continuous air bleed from the pressurization chamber was maintained such that the compressed air sparged into the liquid violently mixed the liquid during the saturation period. After approximately 15 minutes of pressurization, the air bleed and compressed air supply were simultaneously turned off such that the saturated liquid gauge pressure was maintained at 4.6 Kg/sq cm (65 lb/sq in).

The sample volume of sludge to be thickened was immediately placed in the flotation chamber. A valve was then opened allowing a predetermined volume of saturated liquid to quickly flow from the pressurization chamber into the flotation chamber and intimately mix with the sludge. After this operation was complete, timed measurements of sludge interface height were recorded for a period of 15 minutes. During the initial stages of flotation, interface height measurements were recorded every 10 seconds, whereas, measurements at one minute intervals were sufficient near the end of the flotation period.

At the end of the flotation period, samples of float and subnatant were collected. Float was sampled from the float layer produced and subnatant was collected from the sample tap located in the subnatant zone. Total suspended solids were then performed on the samples of feed, float and subnatant.

Data reduction for all flotation tests was in accordance with the procedures outlined in Appendix A. In addition to solids loading rate, the limiting rise rate as calculated from the solids loading rate and bulk separation rate were also determined. All DAF data presented represents predicted, maximum full-scale performance. The scale up factors used were developed from Eimco-PMD experience and are indicated in Appendix A.

Gravity Thickening

All sludges generated during this study were thickened in pilot scale gravity thickeners. Sludge was discharged at 75-100 ℓ/min (20-26 gal/min) from each flocculating clarifier into separate surge tanks on an automatically, timer-controlled basis. Sludge discharge duration was adjusted routinely to maintain a reasonably constant sludge blanket level in each flocculating clarifier. Sludge was pumped at 5.7-7.5 ℓ/min (1.5-2.0 gal/min) for 1.0-2.5 minute durations at 15-30 minute intervals from the surge tanks to each thickener. Specific pumping intervals and durations were adjusted routinely by resetting timers in order to maintain reasonably constant thickener solids loading rates.

Routine daily data collection pertinent to the gravity thickening units involved determinations of feed solids concentrations, overflow and underflow volumes and thickener sludge depth. Underflow and overflow suspended solids concentrations were determined intermittently when the underflow and overflow storage tanks were full.

Solids balances around the thickeners using operating data collected over extended periods of time (several days to several weeks) permitted calculation of average thickener solids loading rate, solids capture and underflow solids concentration.

The approach used to identify pilot thickener performance was to monitor feed and underflow suspended solids concentration, sludge depth and thickener solids loading rate. Thickener performance was identified from periods of stable thickener operation in terms of solids loading rate and sludge depth.

Bench-scale thickening tests were performed routinely in order to expand the thickener performance data base for a variety of thickener operating conditions. Tests were performed using a 2 liter graduated cylinder equipped with a picket rake mechanism, as shown on Figure 6. A plot of sludge interface height versus elapsed settling time was made for each test. Data was then reduced using procedures presented elsewhere.[2]

Vacuum Filter

A typical vacuum filter run involved pumping thickened sludge and lime slurry of known concentrations to the vacuum filter flocculator. The conditioned sludge would flow by gravity from the flocculator to the vacuum filter vat. After starting the filtrate and vacuum pumps, the filter drum would be operated at its lowest speed setting, thereby generally resulting in an easily dischargeable cake. Samples of feed sludge (with and without chemical conditioning), discharged filter cake and filtrate were collected. Data recorded consisted of sludge feed rate, lime slurry feed rate, drum speed, form and dry vacuum

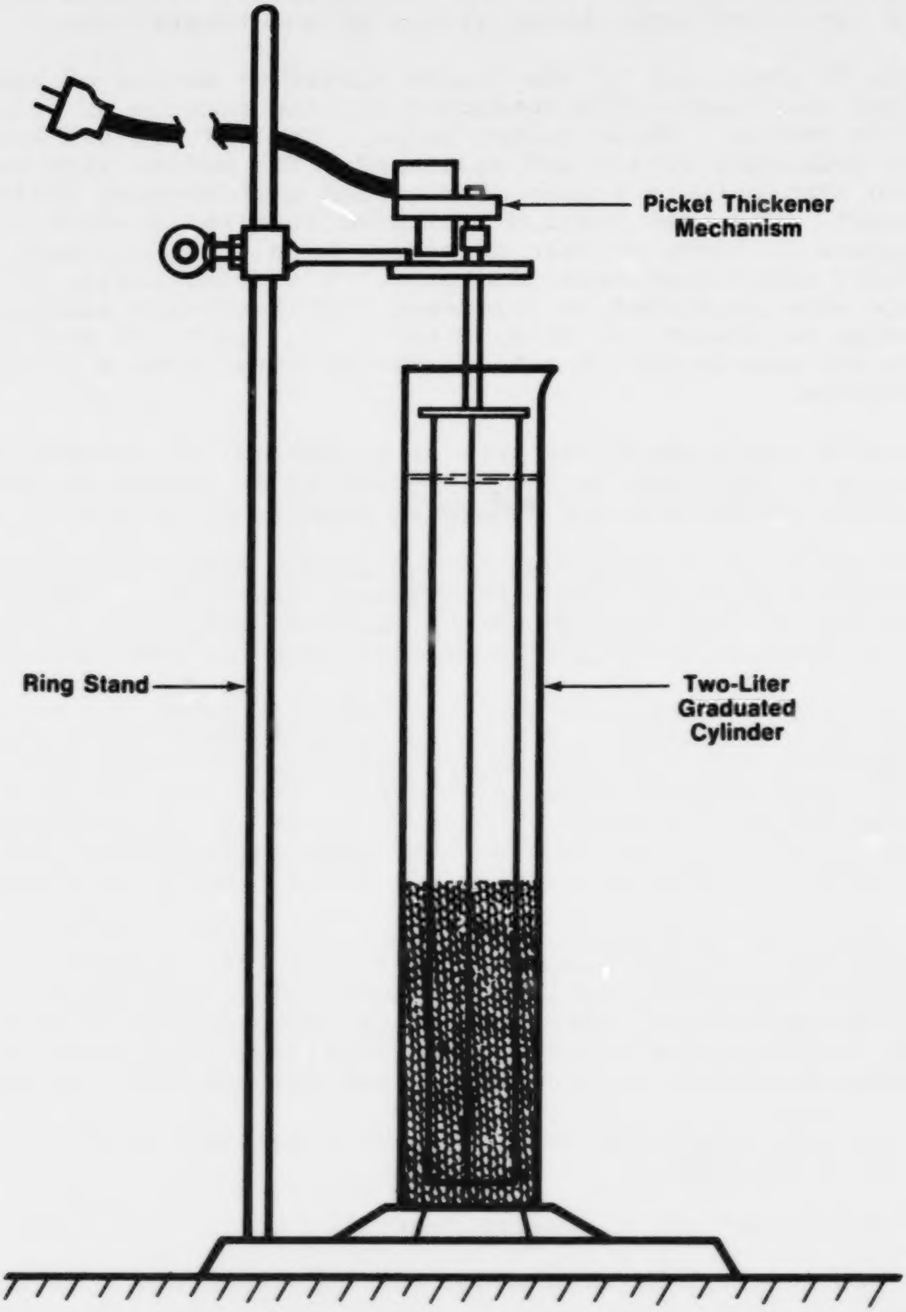
Picket Thickener
Mechanism
Ring Stand
Two-Liter
Graduated
Cylinder

FIGURE 6. SLUDGE THICKENING TEST APPARATUS.

level, cake thickness and dischargeability, filtrate volumetric flow rate, and conditioned sludge pH and temperature.

Prior to operation of the vacuum filter, a series of bench-scale filter leaf tests were conducted for the purpose of selecting a filter medium. Three filter media,* NY-317F, PO-801 and POPR-859, were initially chosen and evaluated. The medias were tested using alum-primary sludge conditioned with several different lime dosages. The test results indicated that the NY-317F media was superior in terms of cake formation rate, cake dryness, filtrate clarity and filter cake discharge. A similar series of leaf tests were performed on thickened ferric-primary sludge. The results indicated the NY-317F media was again the most effective from the standpoint of cake formation rate, cake dryness and discharge.

The drum submergence was kept at a constant 31 percent by maintaining a feed rate to the vat slightly in excess of that capable of being dewatered; the remainder overflowed to drain.

A bridge block arrangement in the vacuum filter valve allowed accurate control of the cake form and dry times. These were arranged so that cake formation occurred during 17.9 percent of one drum revolution with 34.7 percent for cake drying.

During each filter run, cake samples were taken and analyzed for total solids content. To ensure uniform sampling, a round "cookie cutter" with a face area of 77 sq cm (0.08 sq ft) was used. Cake samples were taken after stable operation of the vacuum filter was achieved under a given set of conditions. Usually two or three cake samples were taken during each filter run with the reported cake solids content being an average.

One important observation made during operation of the vacuum filter was a visual judgement of filter cake discharge. The basis for this judgement was a performance code established prior to routine operation. This code established levels of performance with consideration given to cake discharge conditions such as dischargeability, cake cracking, and uniform cake thickness.

Filter cake discharge was judged as excellent, good, fair, poor or not discharging.

An "excellent" (E) discharge meant:

A. The filter cake would readily discharge from the media.

B. Some or frequent cake cracking was observed during the drying portion of the filter cycle.

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*Media designations are those of the Eimco-PMD, Envirotech Corporation

C. Uniform cake dryness was observed between the middle of the sector and the seam.

D. The cake would break into pieces as it passed over the discharge roll.

A "good" (G) discharge meant:

A. The cake would separate and discharge from the media.

B. The cake did not crack during the drying cycle.

C. The observed moisture content was slightly higher in the cake formed on top of the sector seams as compared to the center of the sector.

D. The cake did not crack after going over the discharge roll.

A "fair" (F) discharge meant:

A. The cake would discharge from the media except for occasional timed intervals when a scraper was required to keep the cake discharging.

B. Some filter cake would stick to the seams usually because of a higher moisture content.

C. The cake did not crack after going over the discharge roll.

A "poor" (P) discharge meant:

A. The cake formed on top of the sector seams would not discharge unless aided by a scraper.

B. Cake moisture was visually somewhat higher than the filter cake judged as "fair".

A cake described as not discharging (ND) would have to be scraped off the media.

Centrifuge

The approach used in collecting centrifugal dewatering performance data involved operation at several operating conditions. Feed sludge during any operating day was withdrawn from a single sludge inventory. For each operating condition, samples of feed, centrate, and cake were collected and flowrates of feed and polymer were determined from volume displacements from calibrated feed and polymer tanks. Measurements of bowl rpm and differential rpm were made for each centrifuge run. Solids recovery for each centrifuge run was calculated from a solids balance around the centrifuge. Performance, as outlined above, would be characterized over a range of feed rate, feed concentration, and polymer dose.

An investigation of centrifuge machine variables was performed during previous contract work at this pilot facility on sludges similar to those generated during this study. The results of this previous investigation[7] were used in an effort to minimize the number of machine variables requiring evaluation during this study.

The capacity of larger centrifuges, for the purposes of the economic analysis, was predicted from the performance results developed on the 15.2 cm (6.0 in) diameter machine used during this study. The scale-up factors for solids bowl machines were obtained from the manufacturer.[3]

Filter Press

Prior to each filter press run, sludge conditioning was effected by adding lime on a predetermined dry weight percentage basis. Lime was always added in a slurry form of a known concentration.

Following lime addition to the sludge in the feed tank, the exact volume of conditioned sludge feed was measured. At the end of the run, the final volume of the conditioned sludge was measured which permitted a determination of the amount of sludge used. This served as a cross-check of the techniques used for measurement of filtrate volume generated and the sludge cake volume in the press.

The mechanics of executing each filter press run involved the following established procedures. The feed tank and press were closed and the air compressor started. The pressure regulator was then manually operated to provide a rate of chamber pressure increase of 0.35 Kg/sq cm-min (5.0 lb/sq in-min). After 20 minutes of operation, when a chamber pressure of 7.0 Kg/sq cm (100 lb/sq in) was reached, the pressure regulator was set to maintain this pressure. Data collection during the initial 20 minutes of operation, as well as the remainder of the press run, involved measurement of filtrate flowrate from each chamber. When the filtrate flowrate decreased to 1/20 of the 5 minute filtrate flowrate, the filter press run was terminated. Filter cakes were subsequently removed and analyzed for cake thickness, cake volume, cake wet weight, and cake weight percent total solids. Filtrate, composited from each chamber during the filter press run, was analyzed for suspended solids.

The criteria used in all filter press runs, 0.35 Kg/sq cm-min (5 lb/sq in-min), pressure increase rate to terminal maximum pressure, and filter press run termination at a filtrate rate of 1/20 of the 5 minute rate was arrived at from discussions with various manufacturers of filter press equipment.[5,6]

The media employed in all filter press tests was SA-625* (a saran material). This material was selected from experience with other applications and because of its abrasion resistant properties.

A cloth blinding study was performed on the SA-625 media and indicated it was an acceptable media and did not blind on alum-primary sludge. The study was conducted by performing four consecutive filter press runs on the same sludge (alum-primary in this case) under the same conditions without washing the media between runs. During each run, a plot of form time/filtrate volume versus filtrate volume was made (see Figure 7). The straight line portion of the curve was extended to intercept the time/filtrate volume axis. By using this procedure on each run, the intercept values were compared to determine if cloth blinding was taking place. If blinding was occurring, the intercept would increase in value with progressive filter press runs. From an examination of the intercept values from each run (see Table 2), for both chambers, it was determined that no cloth blinding was occurring.

TABLE 2

SUMMARY OF RESULTS OF FILTER PRESS
CLOTH BLINDING ANALYSIS

	θ/V Intercept $\frac{\text{sec}}{\text{m}\ell}$	
Run Number	Front Chamber	Back Chamber
1	.078	.052
2	.054	.042
3	.058	.046
4	.062	.040
	$\bar{X}$ = .063	$\bar{X}$ = .045

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*Media designations are those of the Eimco-PMD, Envirotech Corporation

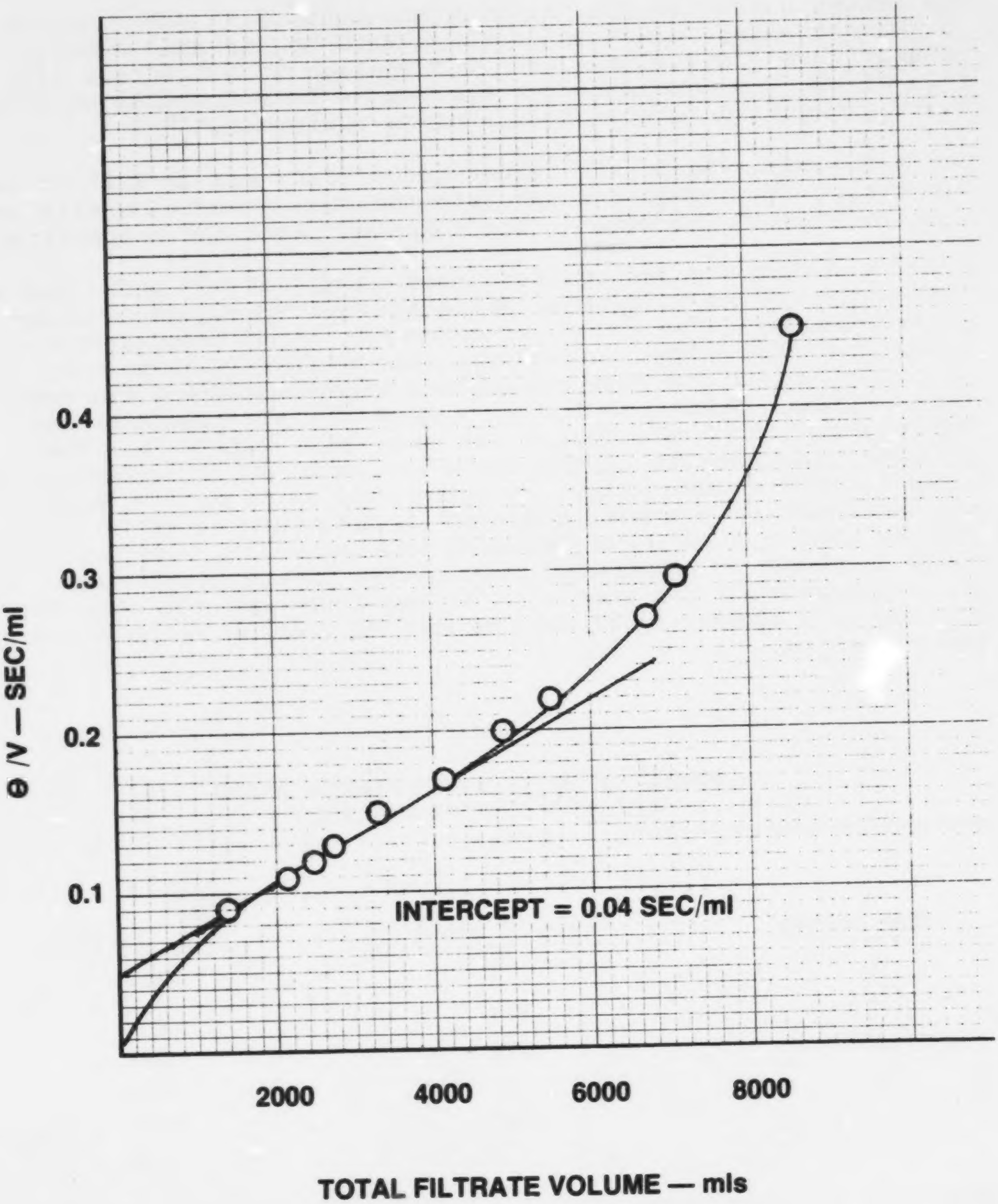

FIGURE 7. SAMPLE DATA PLOT FOR FILTER PRESS CLOTH-BLINDING ANALYSIS.

Specific resistance tests were conducted prior to each filter press test series to determine an acceptable level of lime conditioning (it was learned very early that inadequate sludge conditioning could lead to exorbitantly long form times). Appendix B contains a description of the specific resistance test procedure used.

Figure 8 illustrates the effect of lime dose on specific resistance of alum-primary sludge. Note the rapid drop in specific resistance as the lime dosage increased from 5 percent to 15 percent. Following this rapid drop, a much more gradual decrease occurred between 15 percent and 30 percent.

SAMPLING AND LABORATORY PROCEDURES

Grab samples of alum-primary and ferric-primary sludges from the blowdown surge tanks were collected each weekday. One sample per day was considered acceptable since the underflow blowdown from each CT unit was effectively composited in the 3600 ℓ (950 gallon) blowdown surge tanks.

Once stable operating conditions were established in each chemical treatment unit, sampling of all pertinent wastewater streams was initiated. Beginning each Monday morning and continuing for 48 consecutive hours, composite samples of degritted, screened wastewater and overflow from CT-A and from CT-F were collected. Compositing of the influent was accomplished using an automatic sampler which collected approximately 1 ℓ (0.26 gallon) of sample volume per hour, while the effluent sampler collected approximately 0.1 ℓ (0.026 gallon) of sample volume per hour. Each sample gravity flowed to its respective compositing container which was located in a conventional, household refrigerator maintained at 4°C. After compositing, a representative sample from each container was taken for laboratory analysis of BOD_5, $SBOD_5$, COD, SCOD, turbidity, TSS, total and soluble phosphorus, pH, and methyl orange alkalinity.

In addition to the analyses discussed above, continuous turbidity measurements were automatically recorded for the overflow from each CT unit. These recordings, which were visible in the laboratory, were used by the pilot plant operators for remote evaluation of the operating status of each CT unit. These recordings proved very valuable since they made it possible for pilot plant process performance to be documented at all times, even when the pilot plant was unattended during late night hours.

An analytical laboratory was maintained at the pilot plant site and was sufficiently equipped for the majority of the laboratory analyses performed during this study. All analyses requiring atomic absorption spectrophotometric capabilities were performed in the Instruments Laboratory at Eimco PMD, Envirotech Corporation. Details of specific laboratory analyses performed during this study are discussed in Appendix C.

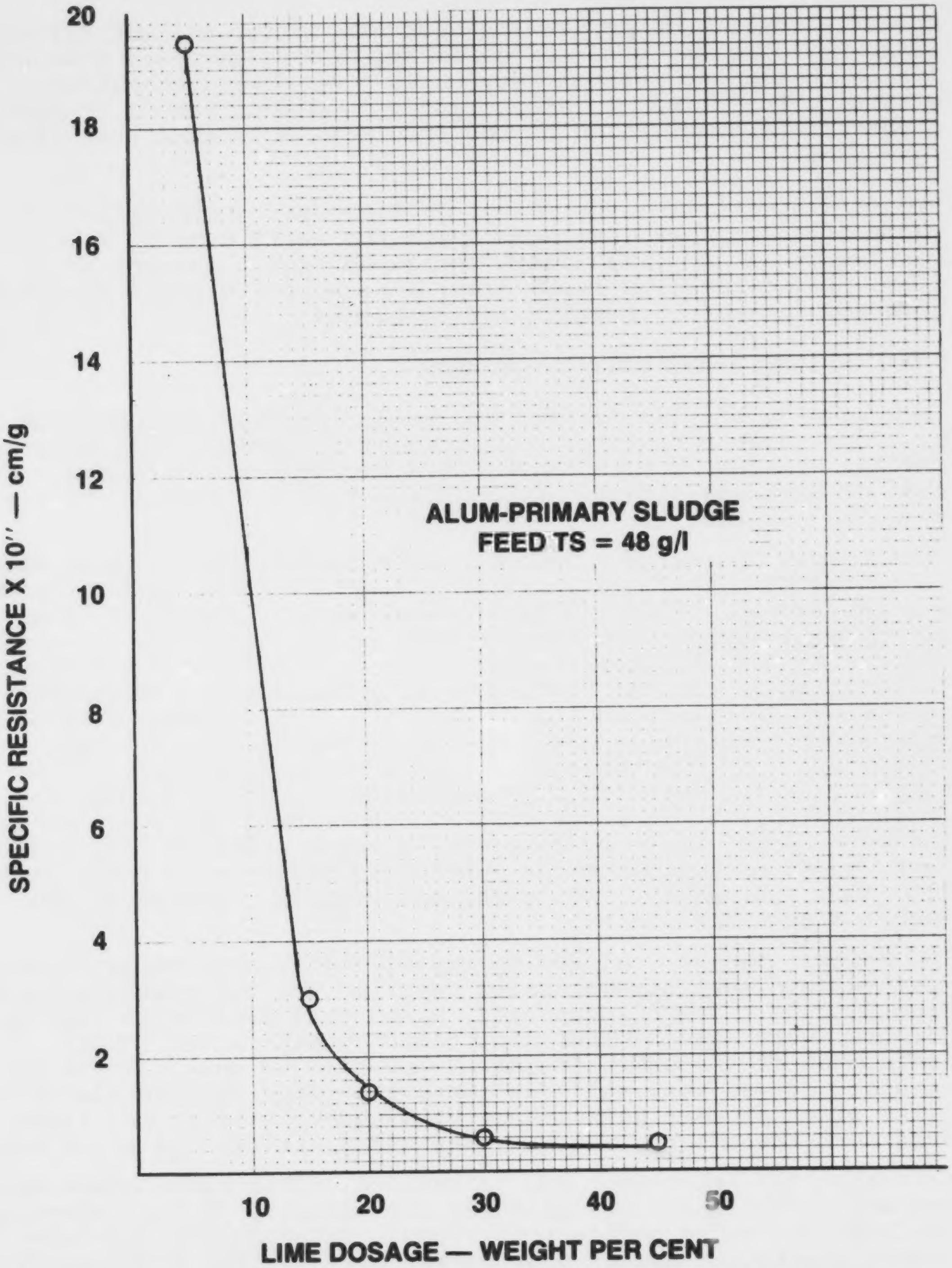

FIGURE 8. EFFECT OF LIME CONDITIONING DOSE ON SLUDGE SPECIFIC RESISTANCE.

SECTION VI

EXPERIMENTAL RESULTS AND DISCUSSION

CHEMICAL TREATMENT

LIQUID PROCESSING SUMMARY

As mentioned, two different chemical dosages were specified to effect two different degrees of phosphorus removal. A high chemical dosage was defined as that dosage required to achieve approximately 95 percent phosphorus removal. The low chemical dosage was defined as the minimum dosage adequate to achieve approximately 80 percent phosphorus removal, or acceptable clarity.

Table 3 shows the average chemical dosage and average polymer dosage for each phase conducted. Also shown is the degritted, screened wastewater flowrate to the flocculating clarifiers and the resulting clarification zone overflow rates. During any phase, any difference in flowrate would arise from a period of downtime in one unit and not the other. When polymer was used, the dosage goal was 0.50 mg/ℓ (this dosage was determined from the jar tests). Note that Phase 4A was essentially identical to Phase 2A; this was the operating goal.

Percent removals were computed for phosphorus, TSS, BOD and COD for each phase and are shown in Tables 4-9. All average figures were calculated from the values determined from each 48 hour composite sample.

From examination of the effluent suspended solids and phosphorus results listed in Tables 4-5, it can be seen that 95 percent phosphorus removal was generally achieved during the high chemical dosage operation. Phase 1F was an exception with a total phosphorus recovery of only 88 percent.

Prior to initiating the low chemical dosage operation of Phase 3, jar tests were conducted to assist in determining chemical dosages required to effect 80 percent phosphorus removal. When these dosages were used in the flocculating clarifiers, however, it was discovered that the ferric chloride dosage initially employed was too high and provided a phosphorus removal greater than that desired. Accordingly, the ferric chloride dosage was decreased after two weeks of operation. The new dosage selected

TABLE 3

SUMMARY OF FLOCCULATING CLARIFIER OPERATING CONDITIONS

Phase*	Average Chemical Dosage (mg/ℓ as Fe/Al)	Average Polymer Dosage (mg/ℓ)	Average Raw Sewage Flowrate (ℓps)	Clarifier Loading (ℓps/sq m)
1A	14.7	-	1.40	0.228
1F	28.3	-	1.39	0.227
2A	15.9	0.46	2.56	0.417
2F	32.8	0.46	2.56	0.417
3A	8.1	0.48	2.47	0.403
3F-I	22.8	0.70	2.44	0.398
3F-II	10.9	0.55	2.47	0.403
4A	15.6	0.49	2.49	0.407

*1A = Phase 1 using alum;
1F = Phase 1 using ferric chloride, etc.

Clarifier loading based on 6.13 sq m clarification area

TABLE 4

CHEMICAL TREATMENT UNIT PERFORMANCE SUMMARY
FROM PHASE 1 OPERATIONS

Value Measured	RAW FEED			CT-A EFFLUENT				CT-F EFFLUENT			
	Average Value	Range	Number of Observations	Average Value	Range	Number of Observations	Percent Removal	Average Value	Range	Number of Observations	Percent Removal
Total BOD_5 mg/ℓ	76	70-81	5	17	12-22	5	78	15	12-20	5	80
Total COD mg/ℓ	88	74-103	5	31	27-37	5	65	27	22-32	5	69
TSS mg/ℓ	66	37-85	5	11	6-15	5	83	(10)34	21-65	5	(85)48
Total Phos. as P mg/ℓ	4.4	4.0-4.8	5	.23	.17-.38	5	95	.55	.27-.76	5	88
Sol. Phos. as P mg/ℓ	3.0	2.2-3.5	5	.06	.008-.19	5	98	.09	.02-.28	5	97
Total $0\text{-}PO_4$ as P mg/ℓ	2.7	2.5-3.4	5	.06	.03-.08	5	98	.30	.11-.52	5	89
Sol. $0\text{-}PO_4$ as P mg/ℓ	2.4	2.1-3.0	5	.012	.006-.026	5	100	.007	<.006-.01	5	100

TABLE 5

CHEMICAL TREATMENT UNIT PERFORMANCE SUMMARY
FROM PHASE 2 OPERATIONS

	RAW FEED			CT-A EFFLUENT				CT-F EFFLUENT			
Value Measured	Average Value	Range	Number of Observations	Average Value	Range	Number of Observations	Percent Removal	Average Value	Range	Number of Observations	Percent Removal
Total BOD_5	81	46-116	7	17	10-26	7	79	14	7-23	7	83
Total COD	89	56-111	7	29	21-40	7	67	27	23-33	7	70
TSS	61	40-95	7	12	7-24	7	80	11	3-19	7	82
Total Phos. P	5.7	4.4-7.3	7	.38	.24-.52	7	93	.35	.15-.57	7	94
Sol. Phos. P	4.1	3.5-.52	7	.22	.05-.50	7	95	.27	.004-.57	7	93
Total $0\text{-}PO_4$ P	3.4	2.3-4.5	7	.08	.04-.13	7	98	.06	.03-.11	7	98
Sol. $0\text{-}PO_4$ P	3.1	2.2-3.8	7	.019	.006-.048	7	99	.013	<.006-.026	7	100

TABLE 6

CHEMICAL TREATMENT UNIT PERFORMANCE SUMMARY FROM PHASE 3A OPERATIONS

Value Measured	RAW FEED			CT-A EFFLUENT			
	Average Value	Range	Number of Observations	Average Value	Range	Number of Observations	Percent Removal
Total BOD_5	138	102-182	8	28	18-37	7	80
Total COD	133	90-163	8	45	36-56	7	66
TSS	108	80-140	8	17	4-35	7	84
Total Phos. P	5.43	4.76-5.97	7	.86	.12-1.32	6	84
Sol. Phos. P	3.52	2.18-4.57	7	.27	.04-.42	6	92
Total 0-PO_4 P	3.46	1.90-4.80	7	.52	.14-.67	7	85
Sol. 0-PO_4 P	3.41	1.68-4.54	7	.20	.02-.32	7	94

TABLE 7

CHEMICAL TREATMENT UNIT PERFORMANCE SUMMARY FROM PHASE 3F-1 OPERATIONS

	RAW FEED			CT-A EFFLUENT			
Value Measured	Average Value	Range	Number of Observations	Average Value	Range	Number of Observations	Percent Removal
Total BOD$_5$	126	105-146	2	21	12-23	2	83
Total COD	114	90-137	2	35	34-35	2	69
TSS	88	80-95	2	10	7-12	2	89
Total Phos. P	5.21	4.76-5.66	2	.27	.21-.33	3	95
Sol. Phos. P	2.49	2.18-2.80	2	.07	.06-.08	2	97
Total 0-PO$_4$ P	3.53	2.53-4.53	2	.09	.07-.10	2	97
Sol. 0-PO$_4$ P	3.53	2.53-4.53	2	.02	.01-.02	2	99

TABLE 8

CHEMICAL TREATMENT UNIT PERFORMANCE SUMMARY FROM PHASE 3F-2 OPERATIONS

Value Measured	RAW FEED			CT-F EFFLUENT			
	Average Value	Range	Number of Observations	Average Value	Range	Number of Observations	Percent Removal
Total BOD_5	142	102-182	6	24	21-26	6	83
Total COD	139	117-163	6	49	40-65	6	65
TSS	115	63-140	6	10	6-19	6	91
Total Phos. P	5.52	4.83-5.97	5	.86	.62-1.07	5	84
Sol. Phos. P	3.94	3.00-4.57	5	.36	.11-.71	5	91
Total 0-P_4 P	3.44	1.90-4.80	5	.50	.29-.65	6	85
Sol. 0-P_4 P	3.36	1.68-4.54	5	.25	.03-.45	6	93

TABLE 9

CHEMICAL TREATMENT UNIT PERFORMANCE SUMMARY
FROM PHASE 4 OPERATIONS

Value Measured	RAW FEED			CT-A EFFLUENT				PRIMARY EFFLUENT			
	Average Value	Range	Number of Observations	Average Value	Range	Number of Observations	Percent Removal	Average Value	Range	Number of Observations	Percent Removal
Total BOD_5	142	90-225	7	24	21-30	7	83	45	39-55	7	68
Total COD	135	119-180	7	43	36-51	7	68	69	60-76	7	49
TSS	120	62-228	7	8	4-17	6	93	32	18-65	7	73
Total Phos. P	5.79	5.13-7.70	7	0.28	.12-.50	7	95	3.99	3.38-4.36	7	21
Sol. Phos. P	4.01	3.53-4.57	7	.07	.03-.20	7	98	3.18	2.66-4.00	7	31
Total $0\text{-}PO_4$ P	3.68	2.80-5.06	7	.09	.03-.17	7	98	2.49	0.84-2.90	7	32
So. $0\text{-}P_4$ P	3.26	1.56-4.40	7	.04	.01-.18	7	99	2.35	0.34-3.13	7	28

reduced the average phosphorus removal to 84 percent, just slightly greater than the 80 percent removal objective. The alum dosage maintained during Phase 3 resulted in a total phosphorus removal of 84 percent. It should be noted that the acceptable effluent clarify criterion was controlling in determining the chemical dosages during Phase 3 operations. A performance summary of Phase 3 chemical treatment operations is shown in Tables 6-8.

During Phase 4, only one of the chemical treatment units was operated to achieve phosphorus removal. Alum was used as the coagulant and the total phosphorus removal goal of 95 percent was achieved. Table 9 summarizes the results of Phase 4 operations.

Discussion of the total suspended solids data shown in Tables 4-9 is warranted. In Table 4 the effluent TSS average concentration from CT-F is shown as 34 mg/ℓ (only a 48 percent removal). As indicated previously, the effluent streams from both flocculating clarifiers were continuously monitored by a recording turbidimeter. Consistently throughout the study, the effluent turbidity continuously recorded from the flocculating clarifier where ferric chloride was being used was lower than that from the alum side of the pilot plant (typically 2-3 versus 5-7 JTU).

A grab sample of the effluent from CT-F was collected and its turbidity measured after several sample storage time periods. During the course of this experiment, it could readily be seen that a post-precipitation effect was taking place as the sample was stored (the test sample was refrigerated to make conditions identical to those of the composited effluent sample). The results of the experiment, shown in Table 10, substantiated the visual observation. It can be deduced from Table 10 that the turbidity of the sample at 48 hours would be approximately 30 JTU. This figure is approximately 10 JTU below the figure actually determined from the 48 composited sample. Since the continuous monitoring turbidimeter indicated a turbidity slightly lower for CT-F effluent than that for the CT-A effluent, it was reasoned that the TSS concentration in CT-F effluent was

TABLE 10

Post-Precipitation Effect in CT-F Effluent

Date: 6-12-74

Time Elapsed (hr)	Sample Turbidity (JTU)
0	1.6
24	15.6
119	56

equal to or less than that in the CT-A effluent. Since no post-precipitation effects were noted in Phases 2 or 3 (due to operation which was aimed at eliminating conditions conducive to long flocculating clarifier sludge retention times) and since the turbidity and TSS concentrations of the CT-F effluent was consistently lower than that of the CT-A effluent (see Tables 5-8 for TSS data), it was estimated that the TSS effluent concentration for CT-F during Phase 1 was 10 mg/ℓ, resulting in an 85 percent removal. These figures are indicated parenthetically in Table 4.

SLUDGE PRODUCTION AND CHARACTERIZATION SUMMARY

Sludge produced in each flocculating clarifier was carefully inventoried so total sludge production could be determined. With the exception of Phase 1, total sludge production for the entire phase was readily computed because all sludge wasted or used for dewatering studies was inventoried.

In Phase 1, it was not possible to use this procedure because inventories and discharge of tanks were not recorded during the entire phase. Therefore, in Phase 1, a representative period of sludge production was taken when it was apparent that steady state operation, in terms of flocculating clarifier sludge blanket level, had been achieved (this representative period for both sludges was plotted and provided a linear slope of sludge production versus elapsed time of operation).

By keeping an accurate inventory of all sludge removed from each flocculating clarifier over the time period of each phase, sludge removal in Kg DS/day could be determined. By knowing the concentration of the effluent suspended solids from each flocculating clarifier (from the weekly 48 hours composited), it was also possible to compute an estimate of the solids lost over the effluent weir. Adding this to the sludge removal calculated from tank inventories gave an estimate for total sludge production from each unit.

An attempt was made to predict total sludge production by devising a model based on influent suspended solids, total phosphorus, and coagulant fed to the flocculating clarifier. The following formula was used to predict total sludge production (including solids lost over the effluent weir) when using alum.

$$\frac{\text{Kg DS}}{\text{day}} = 0.0864\ Q\ \{\text{TSS} + (P)(3.94\ \text{mg Al PO}_4/\text{mg P}) + \text{Al}_r\ (2.89\ \text{mg Al(OH)}_3/\text{mg Al}\} \qquad (1)$$

where:

DS = dry suspended solids

Q = raw feed flowrate, ℓ/sec

TSS = total suspended solids in influent, mg/ℓ

P = total phosphorus in influent, mg/ℓ

Al_r = aluminum available after its utilization in $AlPO_4$, mg/ℓ

The following formula was used to predict total sludge production (including solids lost over the weir) when using ferric chloride:

$$\frac{\text{Kg DS}}{\text{day}} = 0.0864\ Q\ \{TSS + (P)(4.87\ \text{mg}\ FePO_4/\text{mg}\ P) + Fe_r\ (1.89\ \text{mg}\ Fe(OH)_3/\text{mg}\ Fe\} \quad (2)$$

where:

Fe_r = iron available after its utilization in $FePO_4$, mg/ℓ

Assuming that the coagulant fed to the flocculating clarifier combined with the phosphorus as $FePO_4$ or $AlPO_4$, and that excess metal was always fed to the unit, it was assumed, for calculation purposes, that excess metal would exhibit itself as $Fe(OH)_3$ or $Al(OH)_3$.

Table 11 shows the sludge production determined from tank inventories and the predicted sludge production from the above formulas. A summary of the calculations is contained in Appendix E.

TABLE 11

SLUDGE PRODUCTION SUMMARY

Phase	Predicted by Formula (Kg DS/day)	Determined from Inventories (Kg DS/day)
1A	13.9	17.8
1F	15.0	11.5
2A	25.4	33.5
2F	29.2	30.2
3A	29.7	26.9
3F-I	29.1	19.2
3F-II	30.7	26.3
4A	37.4	31.1

Since the indicated formulas take influent suspended solids concentrations, influent phosphorus concentrations, and coagulant dosages fed to the flocculating clarifier into account, it was reasoned that the formula (mass balance) would provide an accurate estimate of the total sludge production. Sludge production as determined from inventories, as listed in Table 11, quantify total sludge production (sludge that was inventoried plus an estimate of solids lost with the effluent). Those solids lost with the effluent, which were based on one 48 hour composite sample per week, represented only 7-8 percent of the total sludge production.

With the exception of Phase 3F-I, reasonable agreement in sludge production determined by formula and inventory was achieved. The maximum deviation in sludge production from the two methods was approximately 30 percent and the average deviation was zero (excluding Phase 3F-I results).

SLUDGE CHARACTERIZATION

Total sludge production was predicted for each type of sludge using mass-balance calculations for each flocculating clarifier (Equations 1 and 2). The sludge produced was further characterized to reflect the amount of chemical solids contained in the chemical-primary sludge. Table 12 lists the sludge characterization for each phase. Percent chemical, as indicated in Table 12 represents the chemical constituents (i.e., $FePO_4$ plus $Fe(OH)_3$) as a percentage of the total sludge dry solids by weight.

Inspection of Table 12 indicates that the volatile fraction of the sludge increased when the chemical dosage decreased. Volatiles for Phases 1 and 2 remained relatively constant. Inerts in Phases 3 and 4 were slightly higher than those in Phases 1 and 2 because the total suspended solids in the raw influent were much higher in the latter two phases. A typical calculation of sludge characterization is given in Appendix F.

TABLE 12

SLUDGE CHARACTERIZATION SUMMARY

Phase	Chemical Dose (mg/ℓ)	TSS_i (mg/ℓ)	P_i (mg/ℓ)	Total Sludge Production (mg/ℓ)	Calculated Balance % Volatiles	Calculated Balance % Chemical*	Calculated Balance % Inert
1A	14.7	66	4.4	115	47	43	10
1F	28.3	66	4.4	125	43	47	10
2A	15.9	61	5.7	115	43	48	9
2F	32.8	61	5.7	132	38	54	8
3A	8.1	108	5.4	139	64	22	14
3F-I	22.8	88	5.2	138	52	36	12
3F-II	10.9	115	5.5	144	65	20	15
4A	15.6	120	5.8	174	56	31	13
4P	-	120	5.8	120	82	0	18

*Expressed as $FePO_4$ + $Fe(OH)_3$ or $Al\ PO_4$ + $Al(OH)_3$ as a percentage of the total DS

DISSOLVED AIR FLOTATION THICKENING

Alum-Primary Sludge

Phase 1 (High Alum; no Polymer)

A summary of alum-primary sludge dissolved air flotation (DAF) thickening results obtained during Phase 1 operations are shown in Table 13 and on Figure 9.

A minimum air to solids weight ratio of approximately 1.4 percent was required for effective flotation thickening to occur. Maximum predicted loading rates were observed to increase with increasing air to solids ratios, whereas, float total solids concentration was insensitive to air to solids ratio above the minimum air to solids ratio necessary to achieve flotation. Total solids capture of 95 percent or higher was achieved at air to solids ratios above the minimum necessary to achieve flotation.

In addition to the successful DAF tests conducted during this phase, the tabulated data summary also includes tests which were classified as marginal or unsuccessful. Tests were rated marginal if flotation occurred but with significant masses of sludge breaking away from the sludge blanket and settling out. Unsuccessful tests were those with operating conditions which resulted in a failure to float the solids.

Phase 2 (High Alum; with Polymer)

A summary of Phase 2 alum-primary sludge DAF thickening results is shown in Table 14 and on Figure 10.

Maximum predicted hydraulic and solids loading rates were found to increase with increasing air to solids ratio while float TSS concentration was insensitive to air to solids ratio. Total solids capture of at least 98 percent was achieved.

Phase 3 (Low Alum; with Polymer)

Table 15 and Figure 11 summarize the DAF thickening results obtained on alum-primary sludge produced during Phase 3.

TABLE 13

DISSOLVED AIR FLOTATION THICKENING OF ALUM-PRIMARY SLUDGE FROM PHASE 1
(High Alum; no Polymer)

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1*	20.3	-	-	0.53	-	-
2*	29.1	-	-	0.56	-	-
3*	29.1	-	-	0.74	-	-
4*	20.3	-	-	0.80	-	-
5*	29.1	-	-	0.94	-	-
6*	11.4	-	-	0.95	-	-
7*	20.3	-	-	1.1	-	-
8	11.4	3.1	216	1.4	7.51	1.33
9	11.4	3.2	246	1.9	8.05	1.80

*Implies marginal or failure to successfully float thicken

Notes

Kg/sq m-hr x 0.204 = lb/hr-sq ft
cu m/sq m-hr x 0.407 = gal/min-sq ft

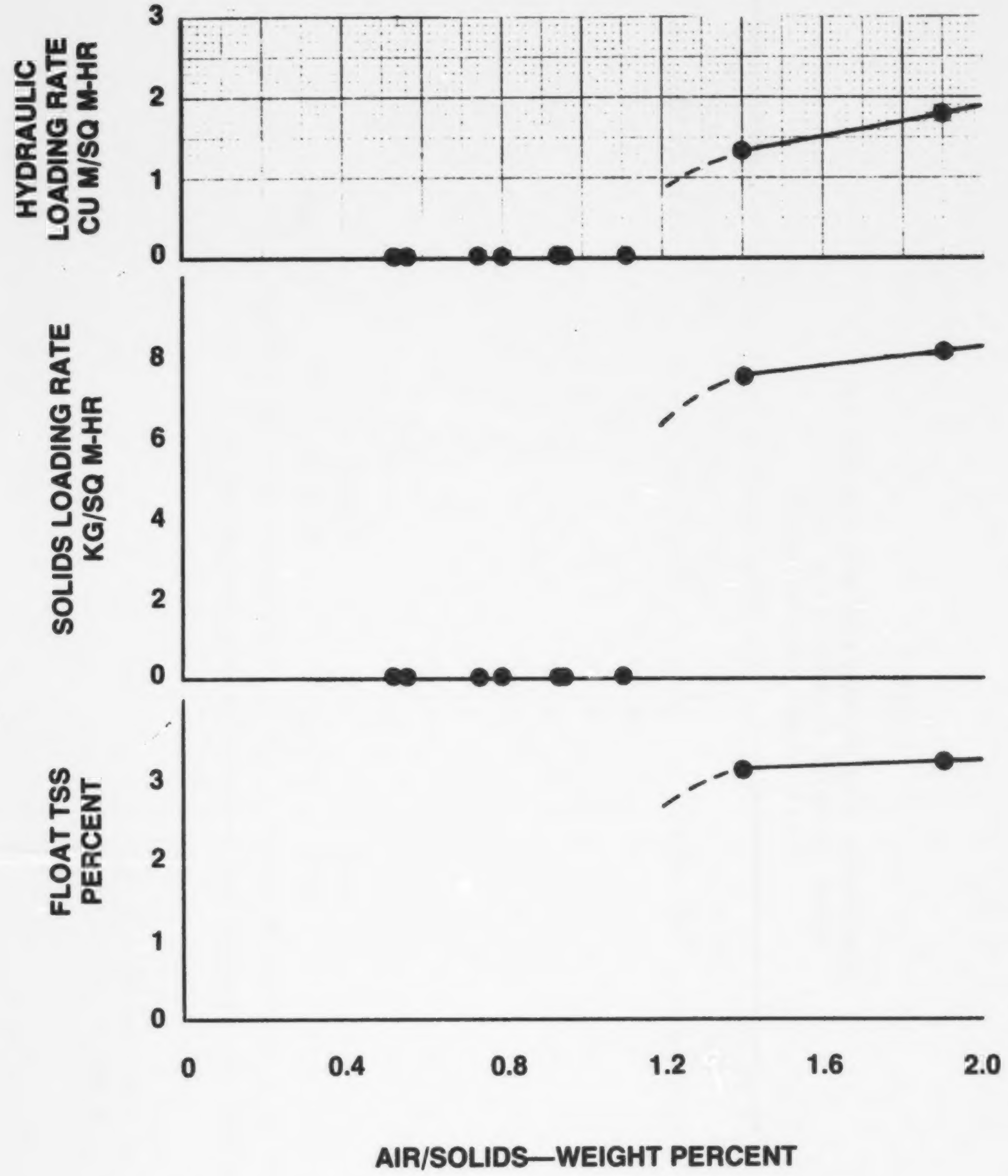

FIGURE 9. DISSOLVED AIR FLOTATION THICKENING OF ALUM-PRIMARY SLUDGE FROM PHASE 1.
(High Alum; no Polymer)

TABLE 14

DISSOLVED AIR FLOTATION THICKENING OF ALUM-PRIMARY SLUDGE FROM PHASE 2
(High Alum; with Polymer)

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1	14.5	2.3	104	0.36	7.55	0.418
2*	8.00	-	-	0.64	-	-
3	8.00	2.3	101	0.64	4.36	0.786
4	14.5	2.3	128	0.71	4.16	0.319
5*	21.5	-	-	0.73	-	-
6	21.5	2.8	60	0.97	6.62	0.909
7	13.8	2.9	84	1.1	4.61	0.835
8	8.00	2.7	116	1.3	7.89	1.97
9	13.8	2.7	70	1.5	5.54	1.18
10	8.00	2.7	132	1.9	6.72	2.09

*Implies marginal or failure to successfully float thicken

Notes
Kg/sq m-hr x 0.204 = lb/hr-sq ft
cu m/sq m-hr x 0.407 = gal/min-sq ft

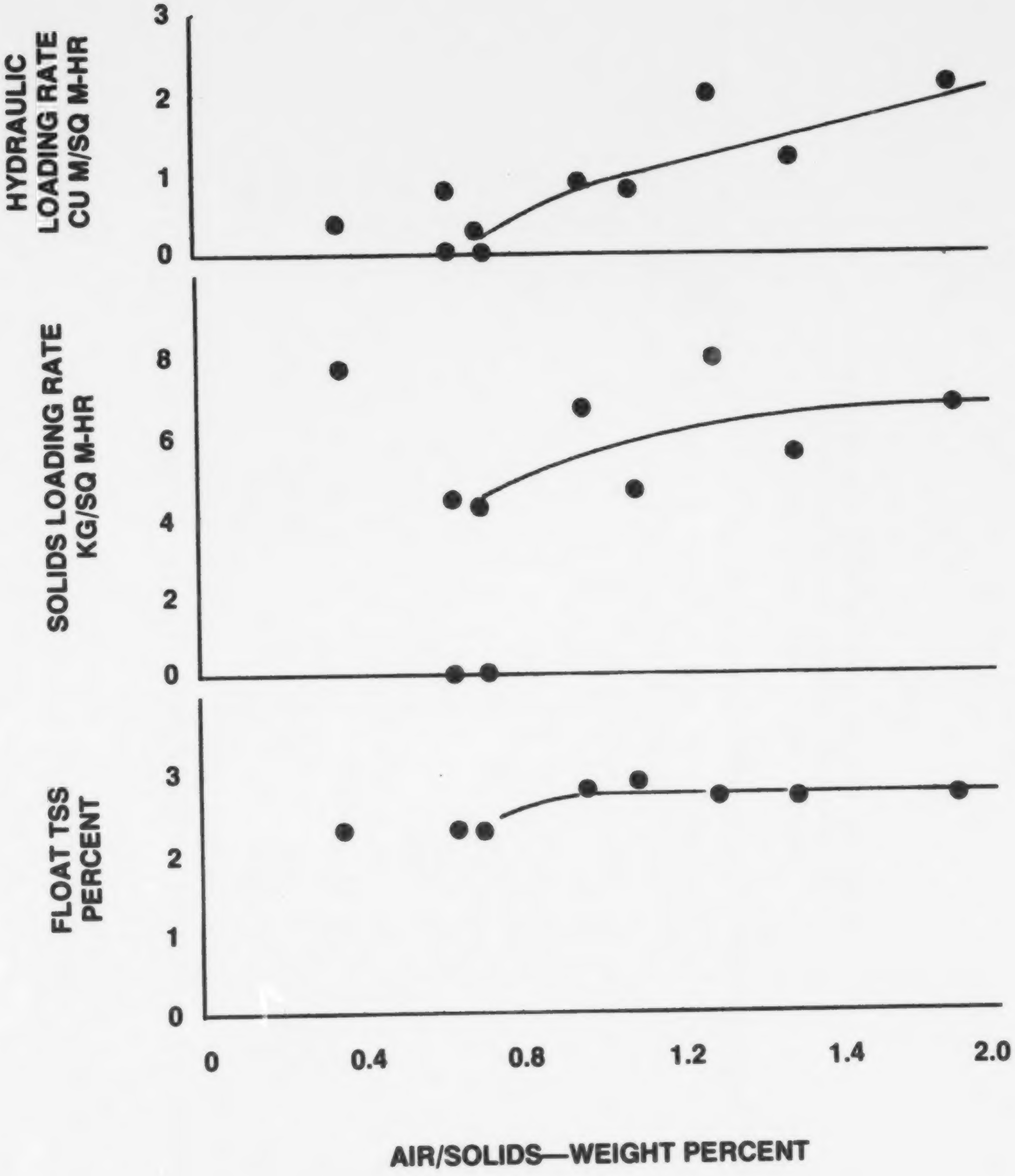

FIGURE 10. DISSOLVED AIR FLOTATION THICKENING OF ALUM-PRIMARY SLUDGE FROM PHASE 2.
(High Alum; with Polymer)

TABLE 15

DISSOLVED AIR FLOTATION THICKENING OF ALUM-PRIMARY SLUDGE FROM PHASE 3
(Low Alum; with Polymer)

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1*	27.8	-	-	0.19	-	-
2	17.5	4.2	306	0.30	6.76	0.580
3	27.8	8.4	424	0.38	7.45	0.400
4	27.8	5.1	410	0.58	9.07	0.816
5	17.5	5.0	314	0.61	10.2	1.16
6	27.8	10.0	480	0.77	16.7	1.80
7	17.5	4.9	250	0.91	14.4	2.04
8	17.5	5.2	96	1.2	24.9	4.32

*Implies marginal or failure to successfully float thicken

Notes

Kg/sq m-hr x 0.204 = lb/hr-sq fr
cu m/sq m-hr x 0.407 = gal/min-sq ft

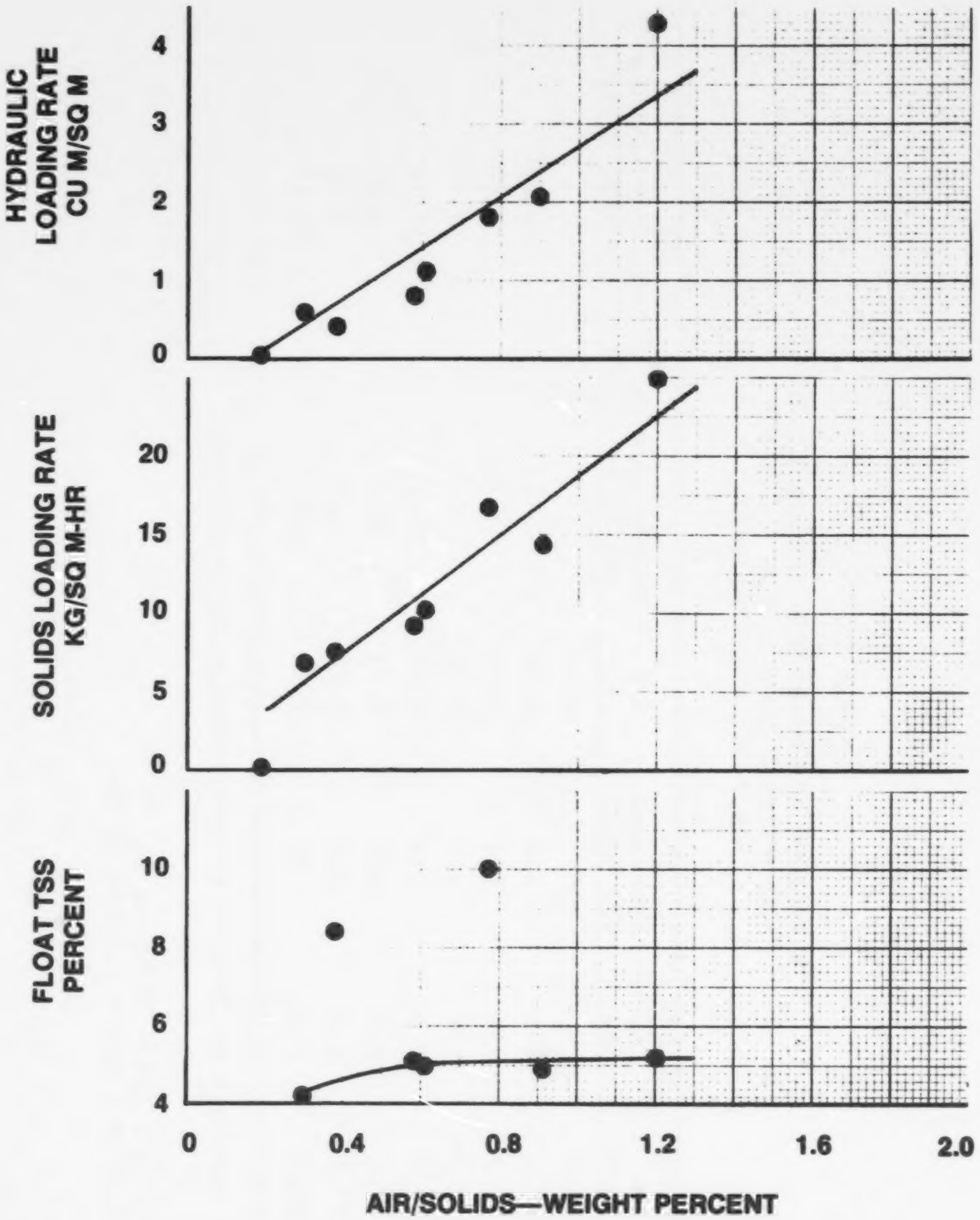

FIGURE 11. DISSOLVED AIR FLOTATION THICKENING OF ALUM-PRIMARY SLUDGE FROM PHASE 3.
(Low Alum; with Polymer)

The effects of air to solids ratio on loading rates and float concentration observed during the first two phases were consistent with those observed during Phase 3. Loading rates and float concentration, however, were higher for the Phase 3 alum-primary sludge. This was reasoned to be due to a lower fraction of chemical solids contained in Phase 3 alum-primary sludge due to the low coagulant dose used during this phase.

FERRIC-PRIMARY SLUDGE

Phase 1 (High Ferric; no Polymer)

Table 16 and Figure 12 summarize the DAF thickening results obtained on ferric-primary sludge produced during Phase 1. Consistent with the trends noted for alum-primary sludge, maximum predicted loading rates increased with increasing air to solids ratio while float concentration was insensitive to air to solids ratio.

Phase 2 (High Ferric; with Polymer)

A summary of the ferric-primary sludge DAF thickening results obtained during Phase 2 is shown in Table 17 and on Figure 13.

Although relatively more scatter was observed in the results from this phase, trends were consistent with previous observations.

Phase 3 (Low Ferric; with Polymer)

Tables 18 and 19 and Figures 14 and 15 summarize the DAF thickening results obtained on ferric-primary sludges produced during Phase 3.

Two distinct ferric-primary sludges were produced during Phase 3. Sludge produced during the initial part of the operating period was generated from operation of the flocculating clarifier with a significantly higher ferric chloride coagulant dose than that used during the latter part. The fraction of chemical solids in the ferric-primary sludge was significantly different, as documented in Section VI.

A review of the DAF thickening results shown on Figures 14 and 15 reveal that the maximum predicted loading rates and float concentration were higher for the Phase 3 ferric-primary sludge containing the lower fraction of chemical solids.

TABLE 16

DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 1
(High Ferric; no Polymer)

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1*	16.5	4.0	246	0.33	8.04	0.737
2*	29.3	-	-	0.37	-	-
3*	41.9	-	-	0.52	-	-
4	29.3	4.8	-	0.55	9.41	0.629
5	16.5	4.8	153	0.66	8.24	1.01
6	29.3	4.8	-	0.73	10.39	1.06
7	16.5	4.8	172	0.99	10.74	1.62

*Implies marginal or failure to successfully float thicken

Notes
Kg/sq m-hr x 0.204 = lb/hr-sq ft
cu m/sq m-hr x 0.407 = gal/min-sq ft

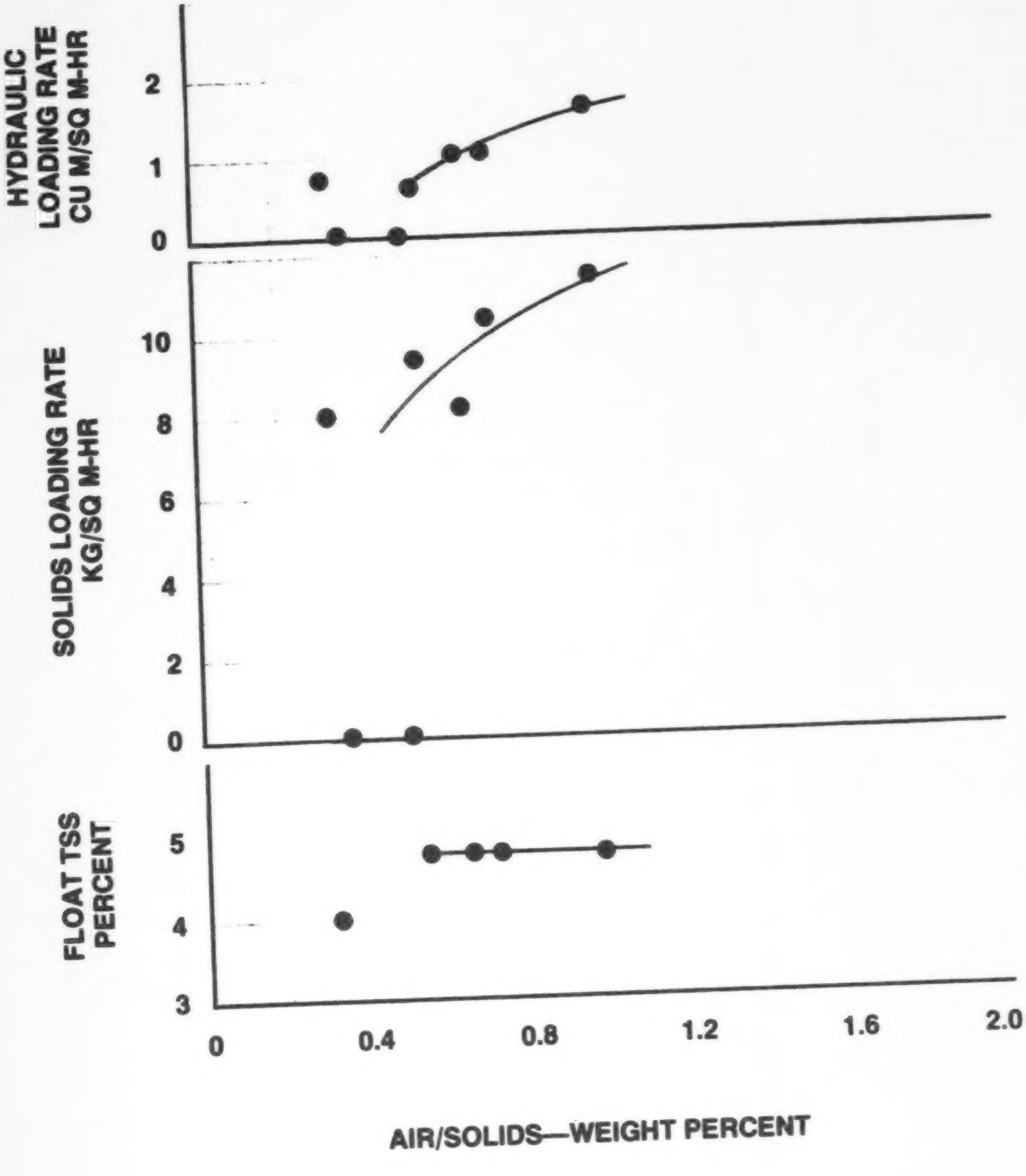

FIGURE 12. DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 1.
(High Ferric; no Polymer)

TABLE 17

DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 2
(High Ferric; with Polymer)

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1*	11.6	3.3	122	0.46	5.83	0.344
2*	20.4	-	-	0.50	-	-
3*	29.1	-	-	0.54	-	-
4*	9.33	-	-	0.56	-	-
5*	29.1	-	-	0.72	-	-
6	20.4	4.0	158	0.77	7.79	0.762
7	11.6	2.8	102	0.93	4.65	0.688
8	20.4	5.1	94	1.0	13.0	1.94
9	11.6	3.4	120	1.3	7.40	0.762
10	11.6	3.5	140	1.8	7.35	1.06
11	11.6	3.7	185	1.8	8.58	2.24

*Implies marginal or failure to successfully float thicken

Notes

Kg/sq m-hr x 0.204 = lb/hr-sq ft

cu m/sq m-hr x 0.407 = gal/min-sq ft

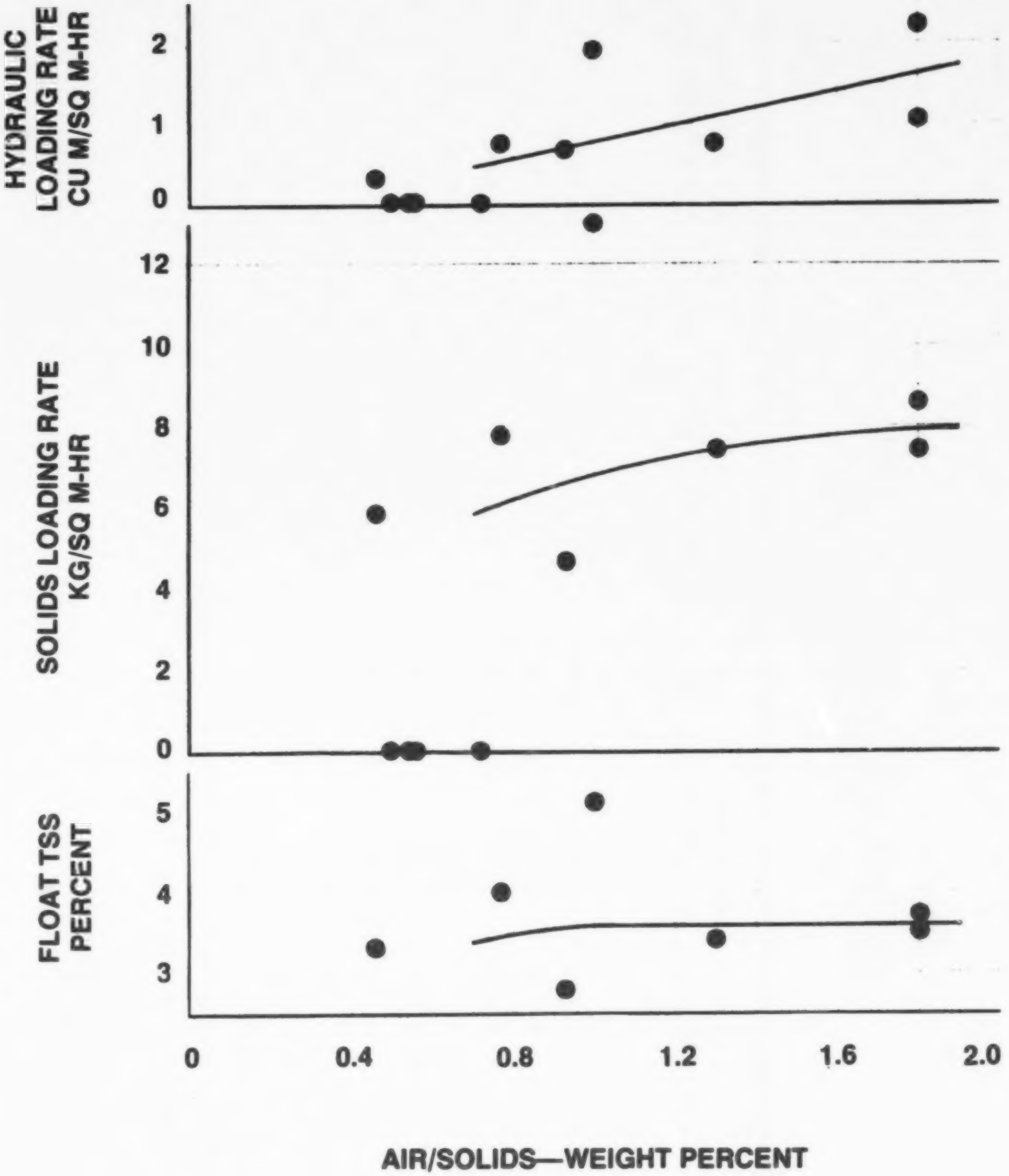

FIGURE 13. DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 2. (High Ferric; with Polymer)

TABLE 18

DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 3-I
(Low Ferric; with Polymer)

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1*	23.5	-	-	0.23	-	-
2*	33.0	-	-	0.32	-	-
3	13.5	3.4	128	0.40	3.97	0.442
4	23.5	3.9	164	0.46	3.92	0.253
5	33.0	4.7	158	0.48	5.10	0.334
6	33.0	4.3	124	0.63	5.54	0.516
7	23.5	4.6	154	0.68	5.05	0.541
8	13.5	3.4	142	0.79	5.78	0.860
9	23.5	4.5	112	0.91	9.22	1.13
10	13.5	4.2	128	1.2	7.30	1.35
11	13.5	3.9	150	1.6	7.94	1.77

*Implies marginal or failure to successfully float thicken

Notes
Kg/sq m-hr x 0.204 = lb/hr-sq ft
cu m/sq m-hr x 0.407 = gal/min-sq ft

TABLE 19

DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 3-II
(Low Ferric; with Polymer)

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1	31.7	4.6	697	0.18	6.37	0.278
2*	22.8	4.3	289	0.24	5.46	0.344
3	31.7	4.9	372	0.35	6.76	0.428
4	22.8	4.0	201	0.50	7.01	0.617
5	31.7	5.0	443	0.53	13.4	1.06
6	31.7	5.3	463	0.70	45.7	2.90
7	22.8	5.2	198	0.76	9.8	1.08
8	13.6	4.0	123	0.79	5.59	0.835
9	22.8	5.1	216	0.94	11.8	1.57
10	13.6	3.9	138	1.2	8.97	1.65
11	13.6	4.9	182	1.6	16.8	3.51

*Implies marginal or failure to successfully float thicken

Notes
Kg/sq m-hr x 0.204 = lb/hr-sq ft
cu m/sq m-hr x 0.407 = gal/min-sq ft

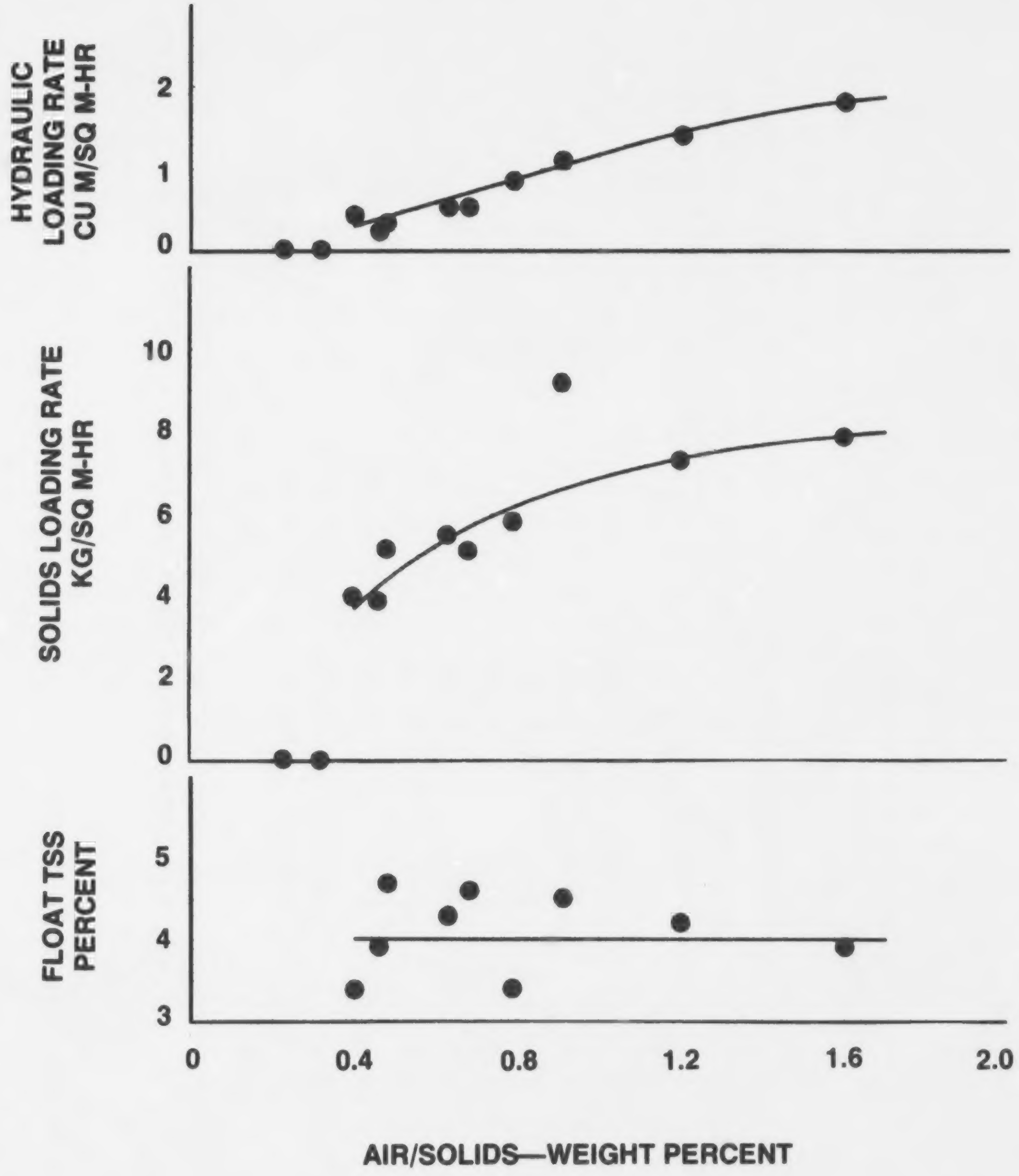

FIGURE 14. DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 3-I.
(Low Ferric; with Polymer)

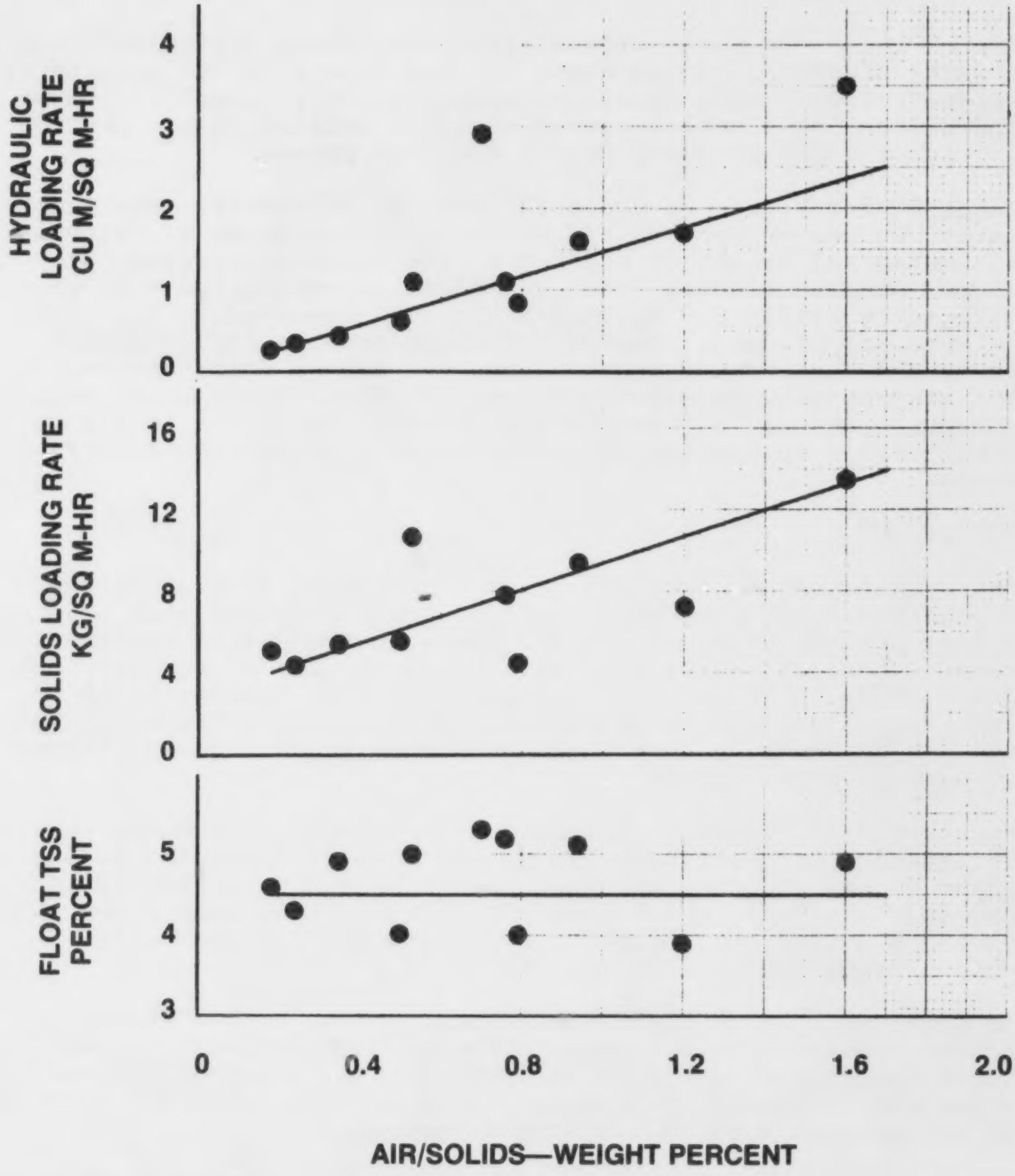

FIGURE 15. DISSOLVED AIR FLOTATION THICKENING OF FERRIC-PRIMARY SLUDGE FROM PHASE 3-II.
(Low Ferric; with Polymer)

PRIMARY SLUDGE

One of the pilot plant flocculating clarifiers was operated as a primary clarifier during Phase 4. The objective was to generate primary sludge (zero percent chemical solids) whose thickening and dewatering characteristics could be related to the chemical-primary sludges produced during previous phases.

Table 20 and Figure 16 summarize the DAF thickening results obtained on the primary sludge produced during Phase 4. Although a substantial amount of scatter in the results occurred, predicted maximum loading rates and float concentration were significantly higher for the primary sludge compared to the chemical-primary sludges. The relationship between air to solids ratio and flotation properties was similar to that observed for alum-primary and ferric-primary sludge. Float total solids content, however, was observed to be more sensitive to air to solids ratio than either the alum-primary or ferric-primary sludge.

DISCUSSION

The weight percent of chemical solids present in alum-primary sludge was correlated with float concentration, maximum predicted solids loading rate and maximum predicted hydraulic loading rate. Chemical solids were defined as the sum of the metal phosphate and metal hydroxide (see Appendix F). Figures 17-19 summarize these correlations. Included in the correlation was the DAF performance results generated on primary sludge (zero percent chemical solids).

Figures 17-19, which include all alum-primary sludge DAF results collected and the primary sludge DAF results collected during Phase 4, were plotted for four levels of air to solids ratio ranging from 0.75 to 2.0 weight percent. The trends shown indicate a worsening in flotation properties with increasing weight fractions of chemical solids in the sludge.

As shown on Figures 17-19, increasing air to solids ratios above approximately 1.5 weight percent had little effect on float concentration or solids loading rate, whereas, maximum allowable hydraulic loading rate increased with further increases in air to solids up to 2.0 weight percent.

Figures 20-22 correlate the weight percent of chemical solids in the ferric-primary sludge versus DAF properties for four ratios of air to solids. The DAF results collected on primary sludge from Phase 4 were included in the summary on Figures in order to extend the analysis to zero percent chemical solids.

TABLE 20

DISSOLVED AIR FLOTATION THICKENING OF PRIMARY SLUDGE FROM PHASE 4

Test No.	Feed TSS (g/ℓ)	Float TSS (%)	Subnatant TSS (mg/ℓ)	A/S (%)	Solids Loading (Kg/sq m-hr)	Hydraulic Loading (cu m/sq m-hr)
1*	27.1	-	-	0.21	-	-
2	16.5	5.3	1800	0.33	55.9	5.06
3	27.1	6.2	604	0.43	59.8	4.42
4	27.1	6.6	640	0.64	40.2	3.69
5	16.5	6.7	1400	0.65	57.4	7.03
6	27.1	7.0	508	0.85	48.0	5.41
7	17.4	7.6	1400	0.93	53.4	7.86
8	17.4	7.5	1200	1.2	67.6	11.8

*Implies marginal or failure to successfully float thicken

Notes

Kg/sq m-hr x 0.204 = lb/hr-sq ft
cu m/sq m-hr x 0.407 = gal/min-sq ft

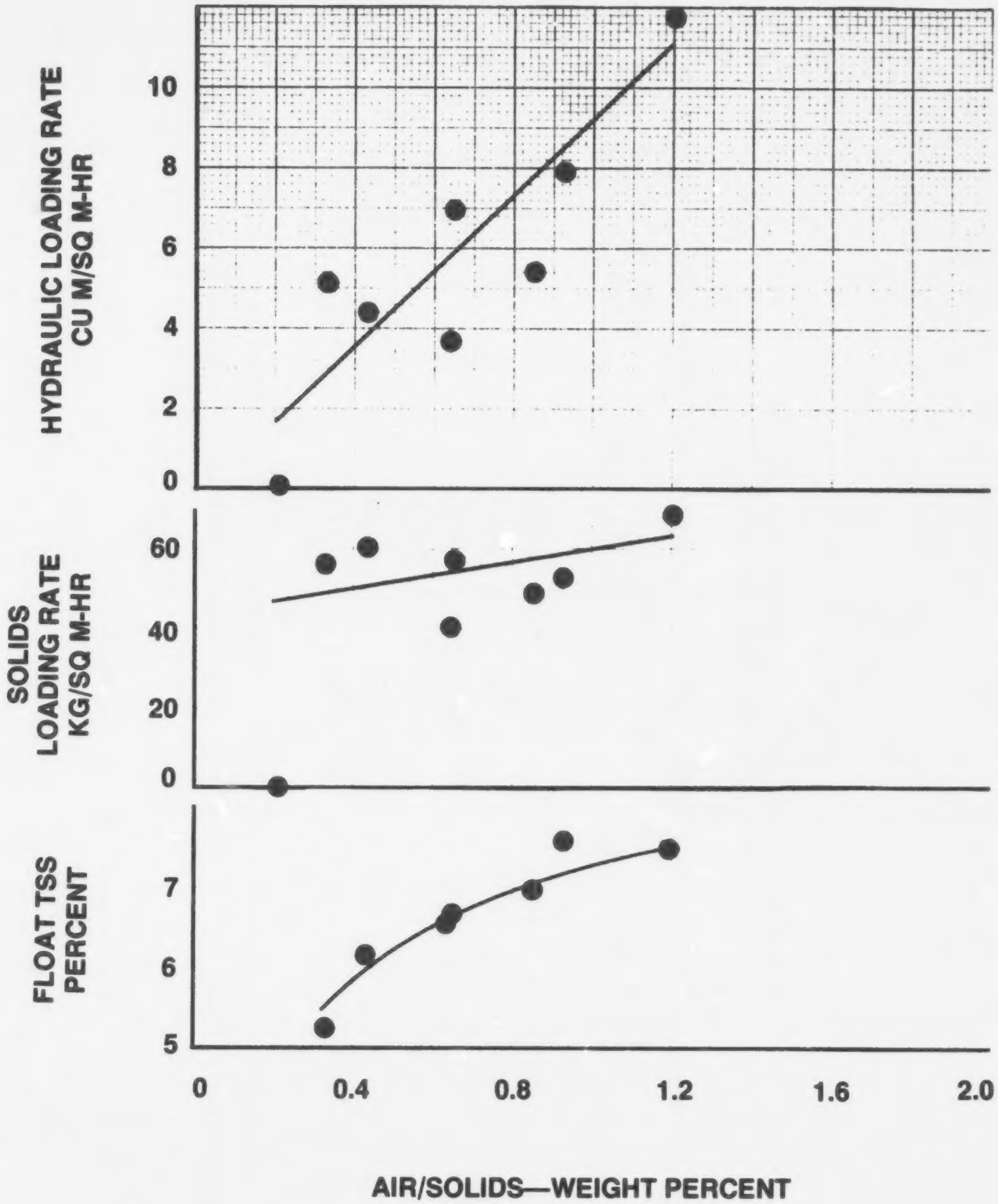

FIGURE 16. DISSOLVED AIR FLOTATION THICKENING OF PRIMARY SLUDGE FROM PHASE 4.

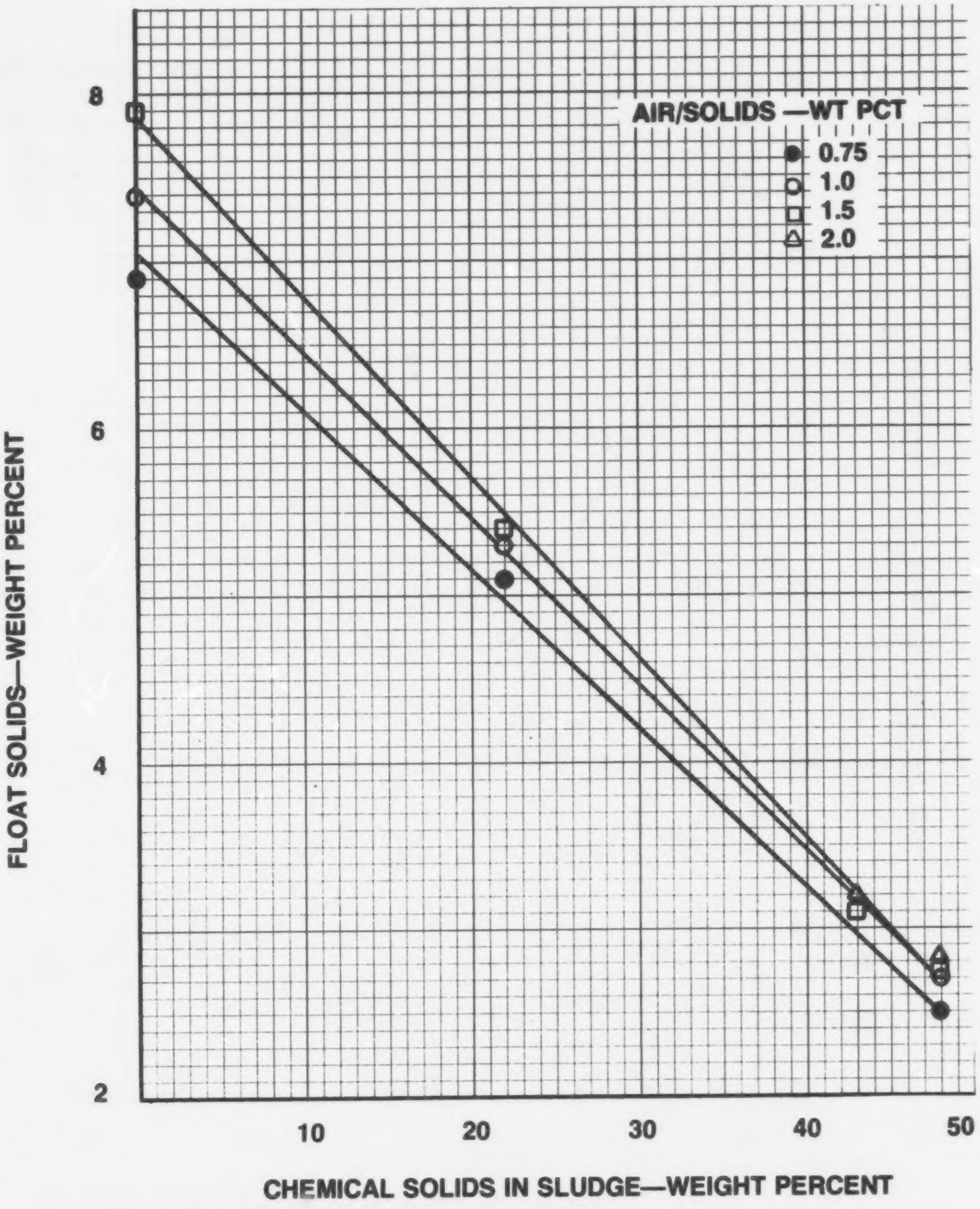

FIGURE 17. FLOAT SOLIDS CONCENTRATION VERSUS FRACTION OF CHEMICAL SOLIDS IN ALUM-PRIMARY SLUDGE.

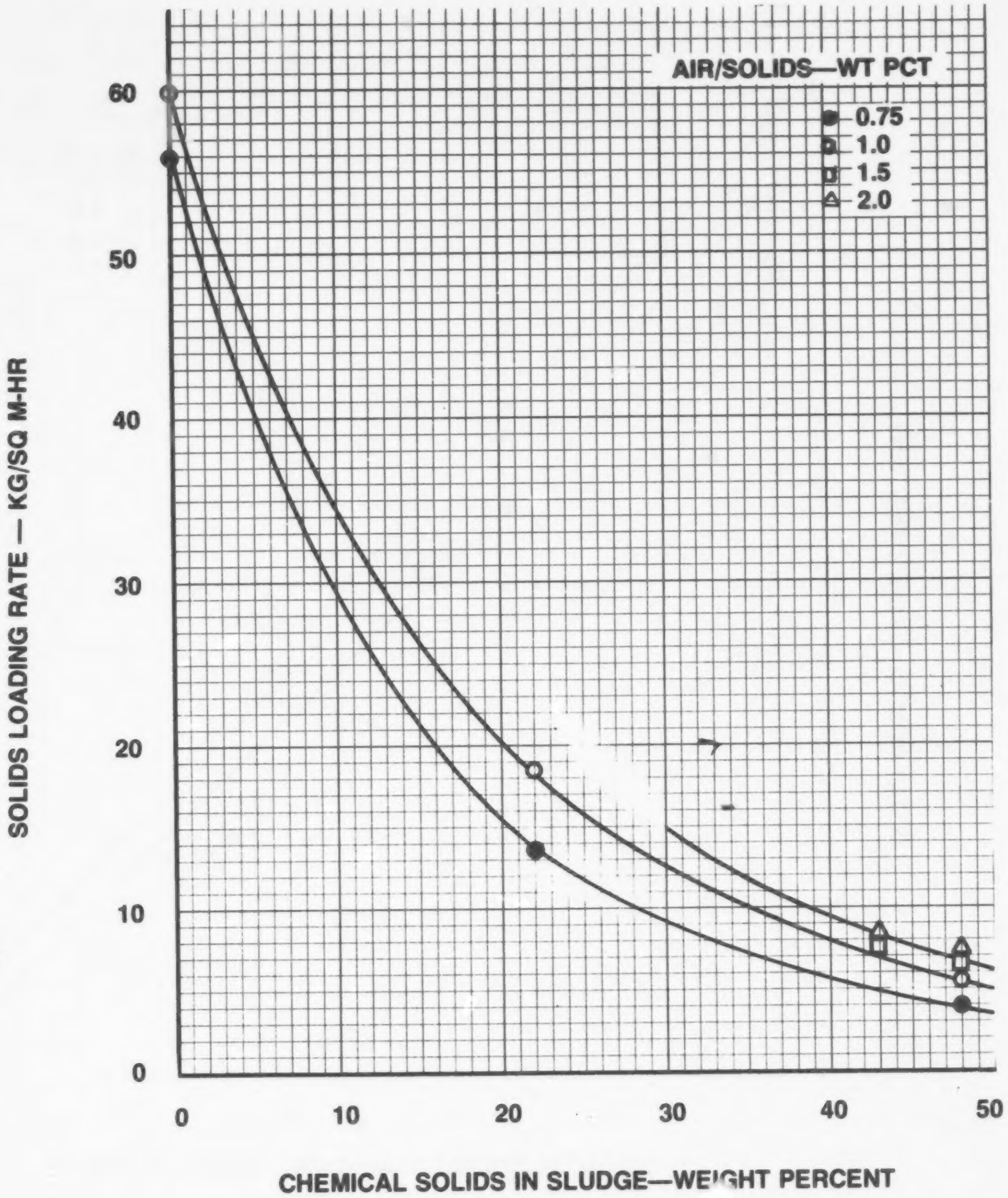

FIGURE 18. SOLIDS LOADING RATE VERSUS FRACTION OF CHEMICAL SOLIDS IN ALUM-PRIMARY SLUDGE.

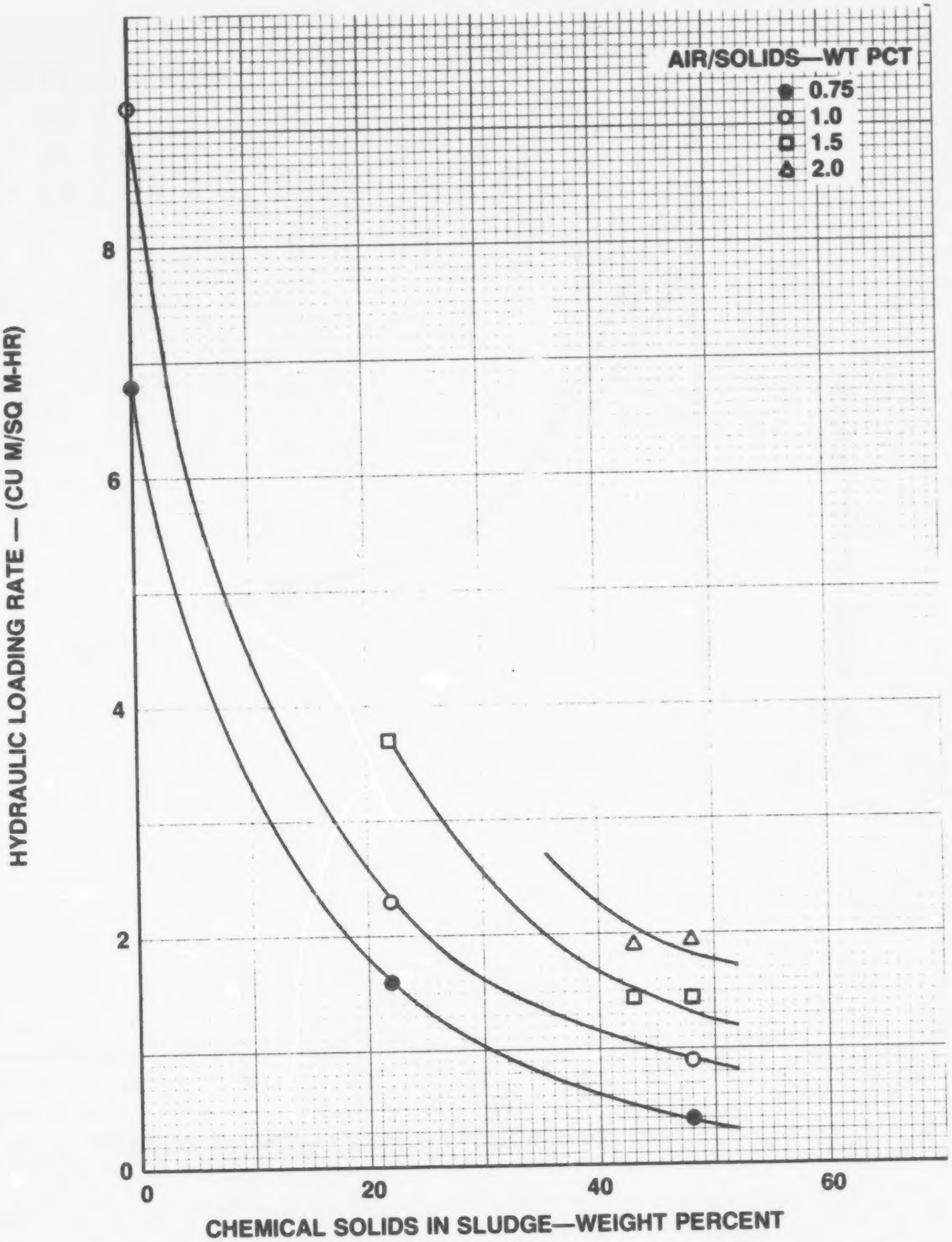

FIGURE 19. HYDRAULIC LOADING RATE VERSUS FRACTION OF CHEMICAL SOLIDS IN ALUM-PRIMARY SLUDGE.

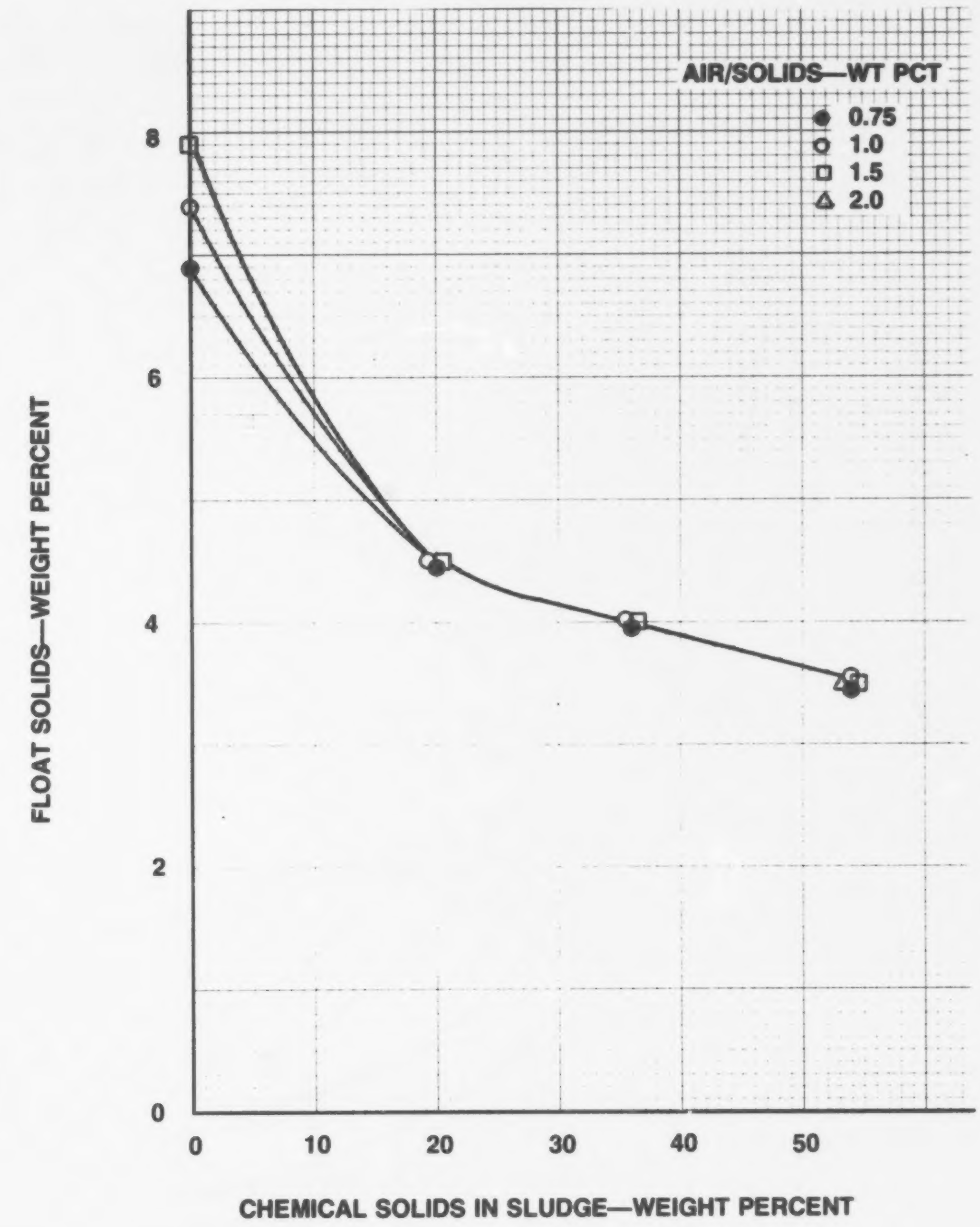

FIGURE 20. FLOAT SOLIDS CONCENTRATION VERSUS FRACTION OF CHEMICAL SOLIDS IN FERRIC-PRIMARY SLUDGE.

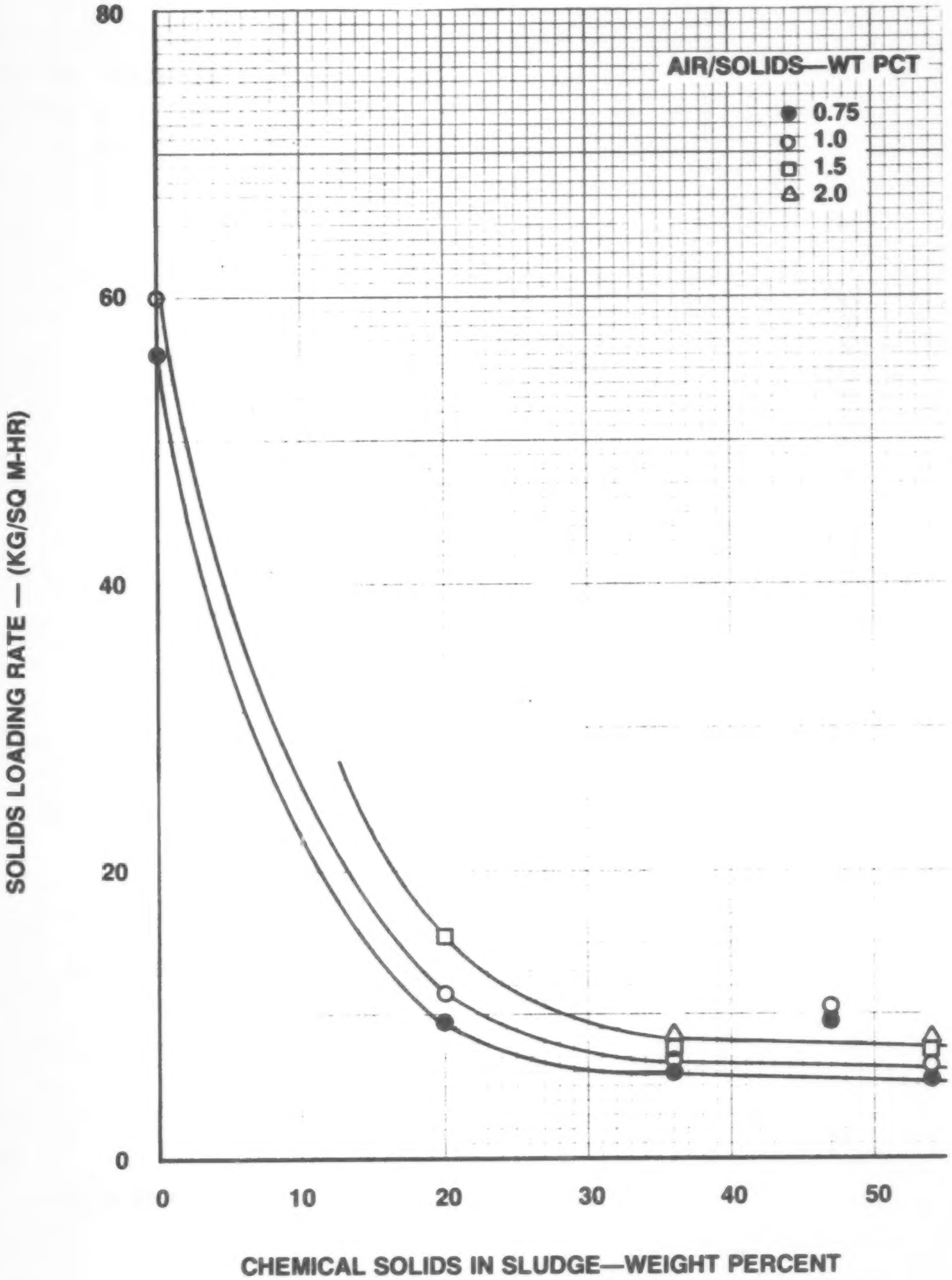

FIGURE 21. SOLIDS LOADING RATE VERSUS FRACTION OF CHEMICAL SOLIDS IN FERRIC-PRIMARY SLUDGE.

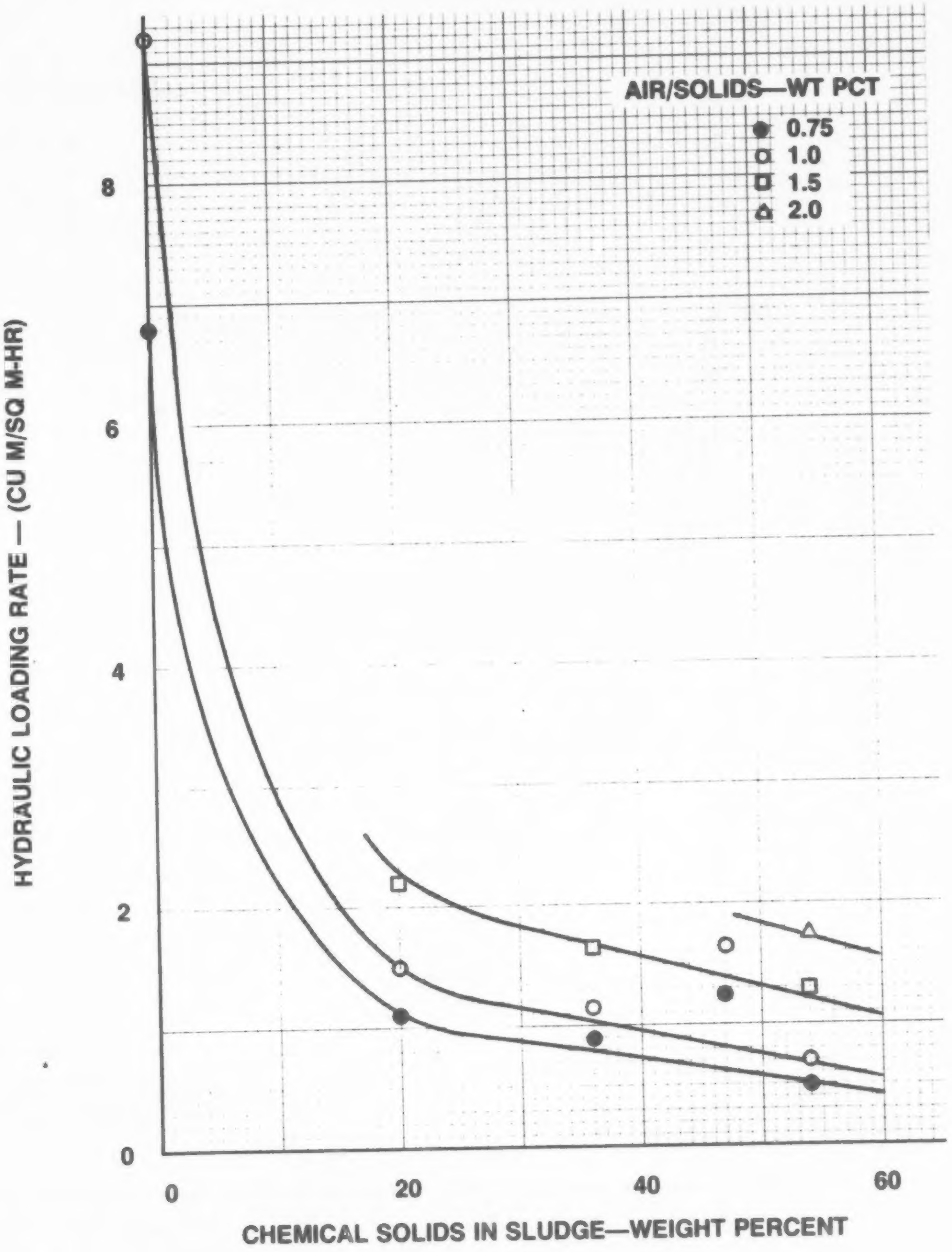

FIGURE 22. HYDRAULIC LOADING RATE VERSUS FRACTION OF CHEMICAL SOLIDS IN FERRIC-PRIMARY SLUDGE

Increased fractions of chemical solids in the sludge generally resulted in poorer ferric-primary sludge DAF properties. Float total solids content was maximized at an air to solids ratio at 0.75 weight percent, and maximum allowable flotator solids loading rate was maximized at an air to solids ratio of about 1.5 weight percent.

From the correlations presented on Figures 17-22, a comparison in DAF properties may be made between the alum-primary and ferric-primary sludges generated during this study. The curves on Figures 23-25 were reproduced in order to facilitate this comparison.

Although an air to solids ratio of 1.5 weight percent resulted in maximization of flotator performance in terms of float concentration and solids loading rate, a ratio of 1.0 weight percent was selected as the basis for the comparison. This was done since a broader and better defined data base was available to an air to solids ratio of 1.0 weight percent as compared to 1.5 weight percent. Inspection of Figures 17-22 revealed that the general outcome of the comparison at a 1.0 weight percent air to solids ratio was similar to that which could be developed at an air to solids ratio of 1.5 weight percent.

Table 21 shows a comparison of DAF performance parameters for alum-primary and ferric-primary sludge at various levels of phosphorus removal. To review DAF performance for these sludges, a raw wastewater containing 100 mg/ℓ TSS and 5.0 mg/ℓ total P was assumed. These characteristics correspond to average Salt Lake City wastewater concentrations observed during this study. The molar ratios of metal to phosphorus required to achieve each level of total phosphorus removal were determined from this study and from previous studies.[1,2]

The results in Table 21 indicates that, for similar levels of phosphorus removal, the DAF performance parameters for alum-primary sludge were superior to those for ferric-primary sludge. The results also indicate a decrease in DAF performance parameters as additional inorganic coagulant is added to achieve higher total phosphorus removal efficiency.

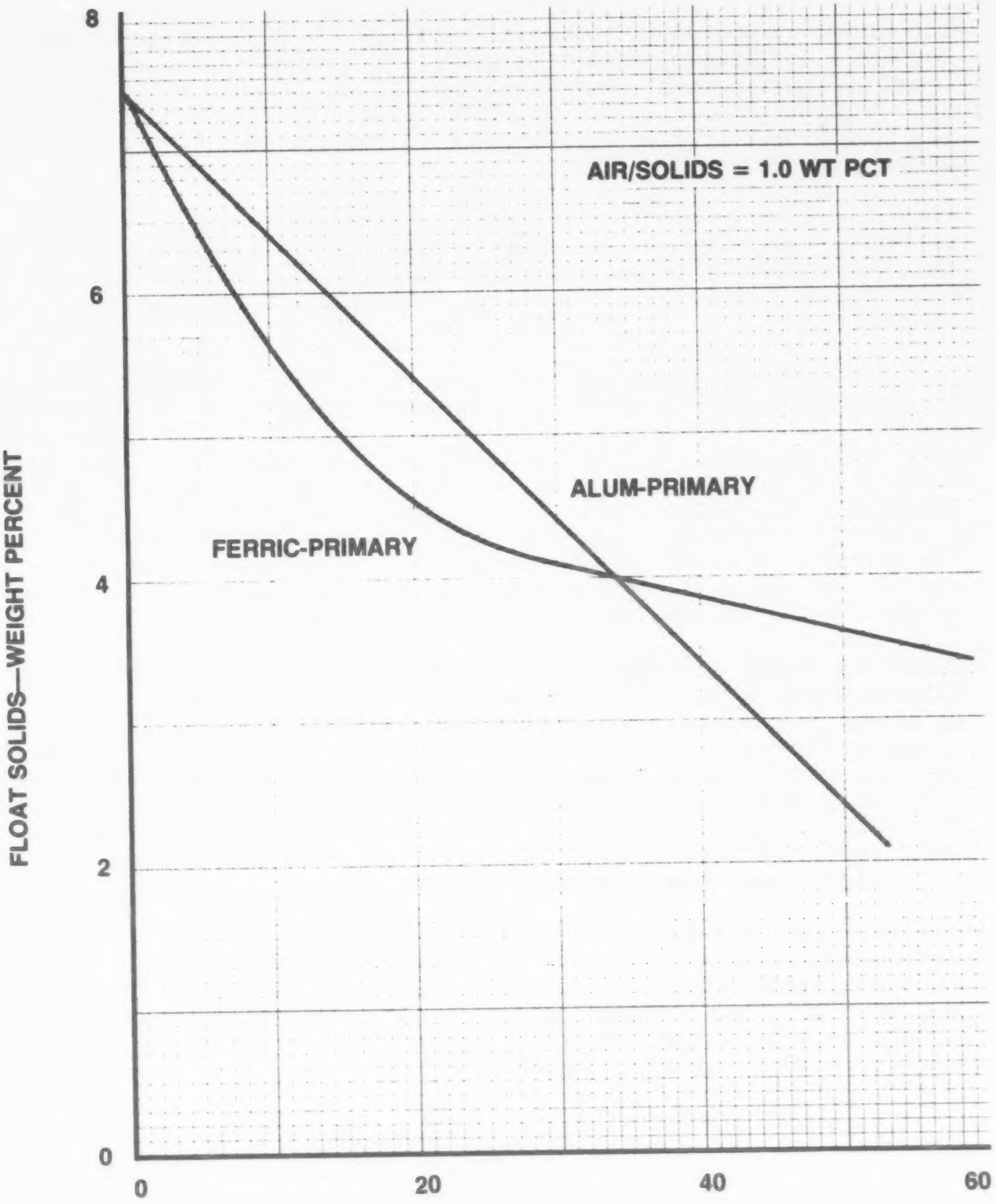

FIGURE 23. COMPARATIVE FLOAT SOLIDS CONCENTRATION OF ALUM-PRIMARY AND FERRIC-PRIMARY SLUDGES.

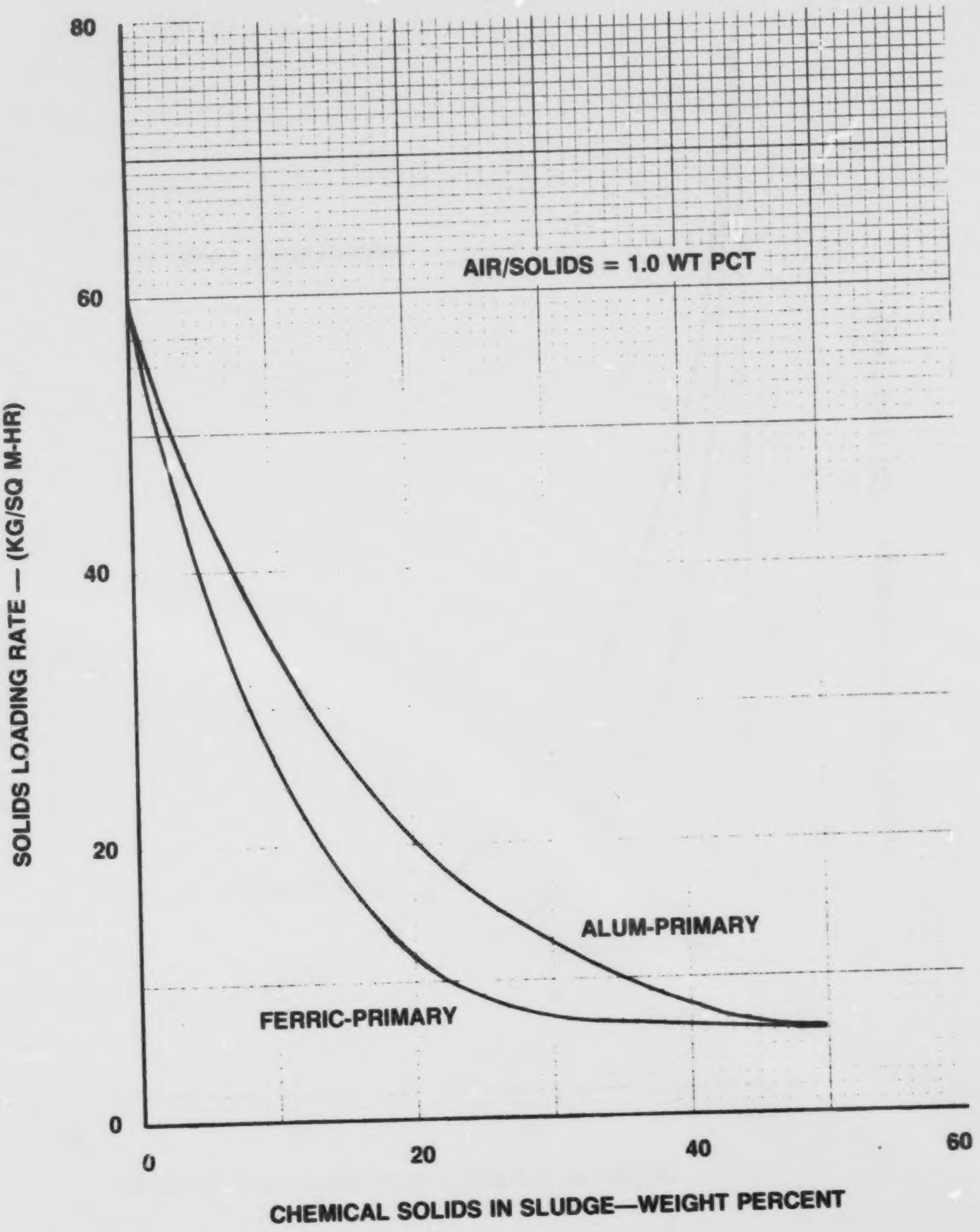

FIGURE 24. COMPARATIVE SOLIDS LOADING RATES OF ALUM-PRIMARY AND FERRIC-PRIMARY SLUDGES.

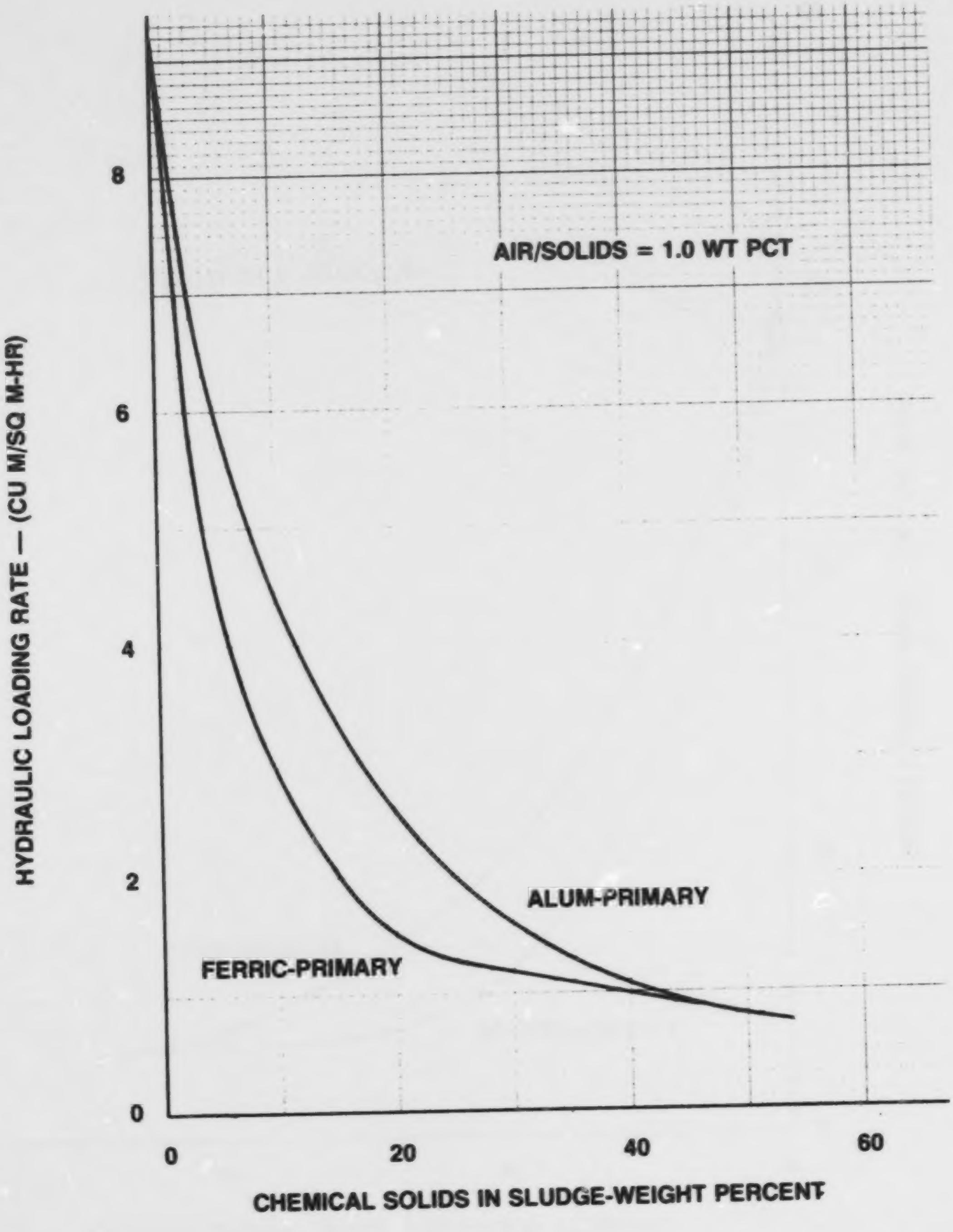

FIGURE 25. COMPARATIVE HYDRAULIC LOADING RATE OF ALUM-PRIMARY AND FERRIC-PRIMARY SLUDGES.

TABLE 21

EFFECT OF PHOSPHORUS REMOVAL ON DISSOLVED AIR FLOTATION PROPERTIES OF ALUM-PRIMARY AND FERRIC-PRIMARY SLUDGE

Total Phosphorus Removal-Percent	Alum-Primary Sludge				Ferric-Primary Sludge			
	Chemical Sludge Weight Percent	Float Percent TS	SLR Kg/sq-m hr	HLR cu m/ sq m-hr	Chemical Sludge Weight Percent	Float Percent TS	SLR Kg/sq m hr	HLR cu m/ sq m-hr
80	18	5.6	24	3.0	22	4.4	9.0	1.6
90	23	5.1	18	2.3	28	4.2	7.5	1.3
95	32	4.2	11	1.4	38	3.9	6.8	1.0

BASIS: Raw wastewater with 100 mg/ℓ TSS and total phosphorus of 5 mg/ℓ as P.
DAF operation at 1% air to solids ratio.

GRAVITY THICKENING

ALUM-PRIMARY SLUDGE

Phase 1 (High Alum; no Polymer)

A summary of gravity thickener performance for alum-primary sludge during Phase 1 is shown in Table 22. As shown in Table 22, three separate stable operating periods were obtained.

Feed concentration ranged from 21-26 g/ℓ, with corresponding underflow concentrations ranging from 23-29 g/ℓ. Clearly, the alum-primary sludge generated during Phase 1 did not gravity thicken to any appreciable extent. Underflow concentration was not affected by solids loading rate over the range studied (at a sludge age of 1.7-2.2 days and sludge depth of 0.95-1.2 m {3.1-3.9 ft}). The only trend discernable was that as the feed concentration increased, the underflow concentration also increased.

The effect of sludge depth within the thickener was studied by sampling the sludge blanket at intervals of 15.2 cm (6 inches) and analyzing these samples for total solids. The results of three typical thickener profiles are shown on Figure 26. These data indicate that a thickener sludge blanket depth of approximately 40 cm (1.3 ft) was required to achieve maximum underflow concentration.

Figures 27 and 28 summarize the results of bench-scale thickening tests performed on alum-primary sludge generated during Phase 1. Plotted on Figure 27 are the pilot plant thickener performance results presented in Table 22. None of the three pilot plant operating periods produced an underflow concentration equal to that predicted by the laboratory tests. Highest underflow concentration from the pilot plant thickener was 2.9%, on Figure 27, this operating point appears in an area where the bench-scale tests predicted an underflow concentration in excess of 5.0%. Figure 28 shows the positive relation between thickener feed concentration and maximum predicted underflow concentration; this trend is substantiated by the pilot plant thickener results.

Phase 2 (High Alum; with Polymer)

A gravity thickener performance summary for alum-primary sludge generated during Phase 2 is shown in Table 23. As shown in Table 23, three stable operating periods were obtained.

TABLE 22

ALUM-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 1 PILOT PLANT RESULTS
(High Alum; no Polymer)

Parameter	Values		
Evaluation period, days	5	12	6
Thickener feed, g/ℓ	21	22	26
Thickener loading, ℓ/min-sq m	.53	.32	.45
Thickener loading, Kg/day-sq m	15.8	10.1	16.8
Thickener sludge age, days	1.7	2.2	1.9
Thickener sludge depth, m	1.2	.95	1.1
Solids capture, percent	98	99	99
Thickener underflow sludge, g/ℓ	23	24	29

TABLE 23

ALUM-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 2 PILOT PLANT RESULTS
(High Alum; with Polymer)

Parameter	Values		
Evaluation period, days	9	6	9
Thickener feed, g/ℓ	21	25	21
Thickener loading, ℓ/min-sq m	.53	.51	.68
Thickener loading, Kg/day-sq m	15.9	18.1	20.9
Thickener sludge age, days	1.9	1.5	1.0
Thickener sludge depth, m	1.2	1.0	.88
Solids capture, percent	99	99	99
Thickener underflow sludge, g/ℓ	27	30	25

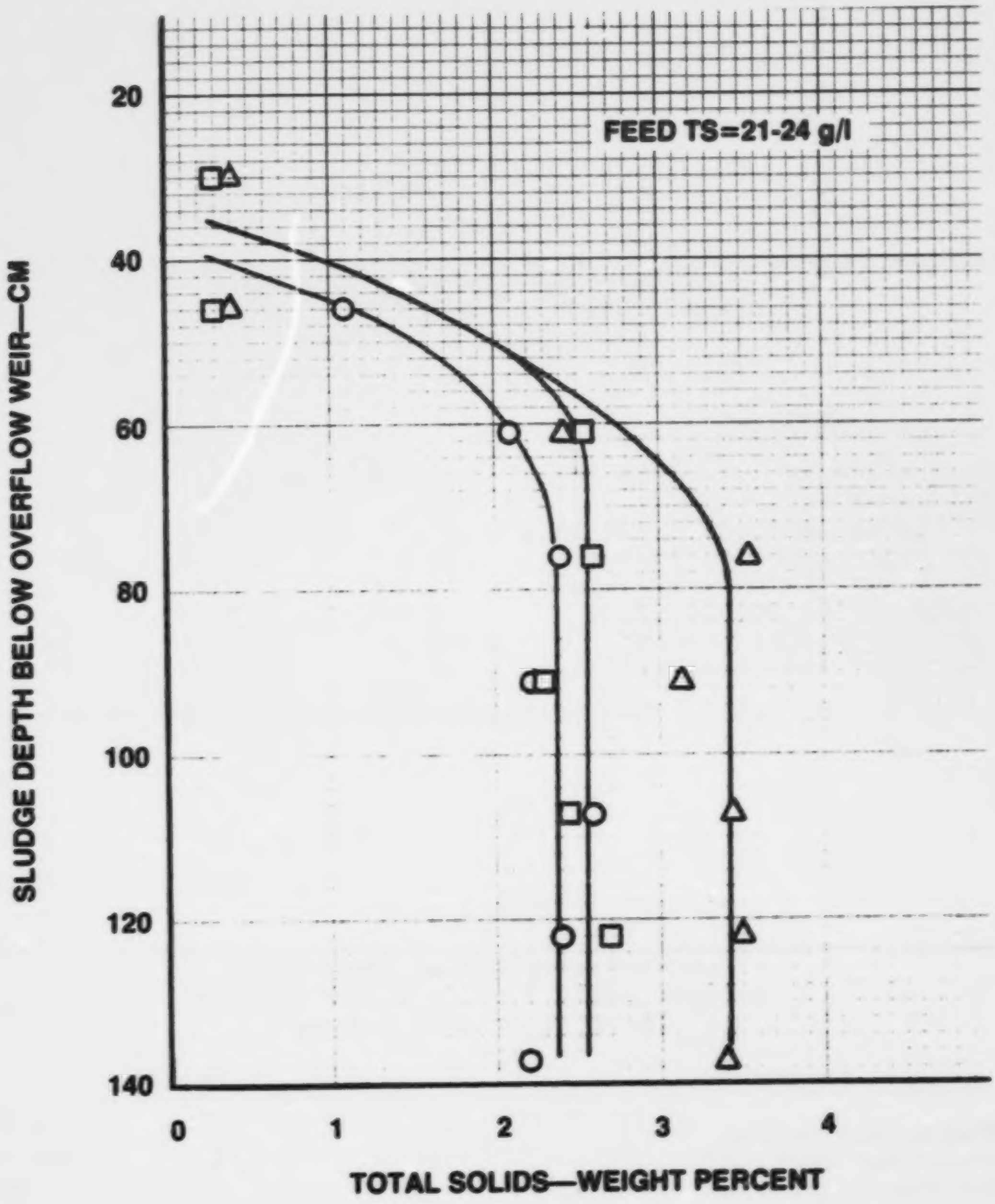

FIGURE 26. ALUM-PRIMARY SLUDGE THICKENER PROFILES PHASE I.
(High Alum; no Polymer)

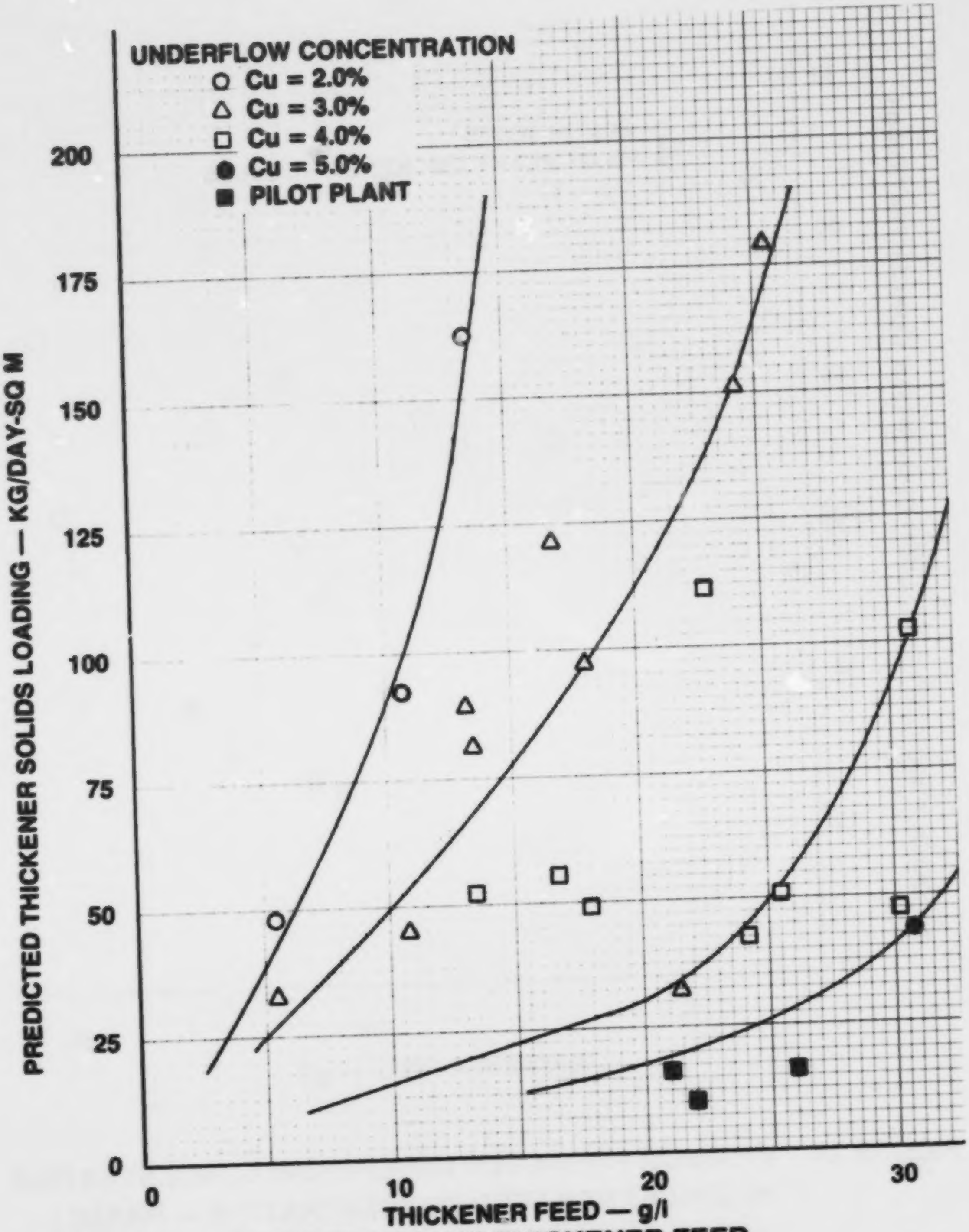

FIGURE 27. PREDICTED TSL VS THICKENER FEED FOR ALUM-PRIMARY SLUDGE — PHASE I. (High Alum; no Polymer)

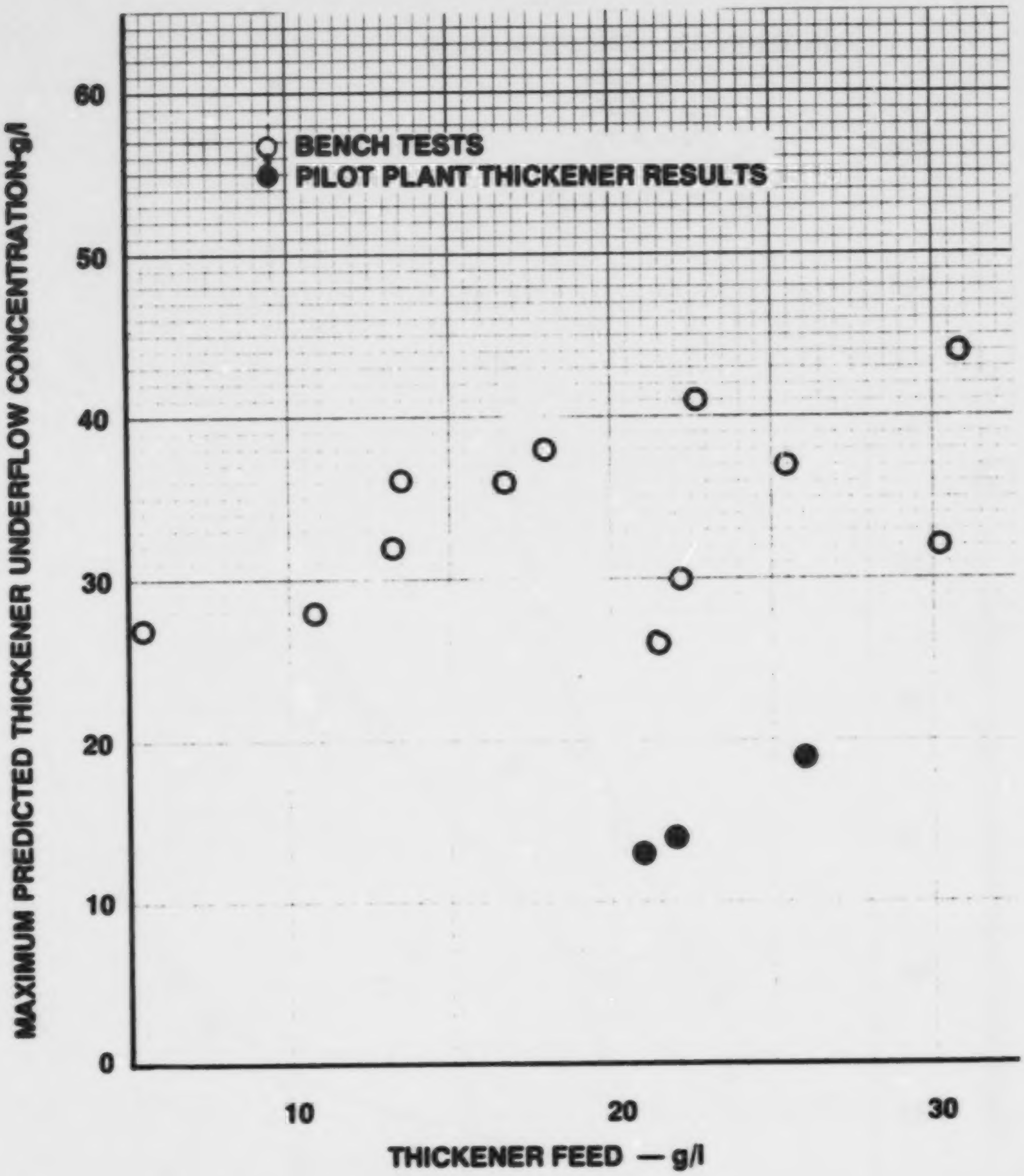

FIGURE 28. MAXIMUM PREDICTED UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION — PHASE I. (High Alum; no Polymer)

Feed solids concentration ranged from 21-25 g/ℓ, and the resulting underflow solids ranged from 25-30 g/ℓ. Solids loading rate, in the range studied, did not affect the underflow solids (at a sludge age of 1.0-1.9 days and sludge depth of 0.88-1.2 m (2.9-3.9 ft). Solids capture was excellent, with all operating periods exhibiting 99% capture. Data given in Table 23 tend to show that a higher feed solids concentration resulted in a thicker underflow concentration.

By conducting thickener profile tests, the effect of sludge depth within the thickener on underflow solids concentration was studied. The sludge blanket was sampled at 15.2 cm (6 inch) depths and these samples analyzed for total solids concentration. Results of three typical tests are shown on Figure 29. All profiles indicate that a sludge depth of only about 20 cm (0.66 ft) was required to achieve maximum underflow concentration. Sludge depths greater than 20 cm (0.66 ft) were of little value in increasing the underflow solids concentration.

The results of bench-scale thickening tests conducted on the alum-primary sludge generated during Phase 2 are summarized on Figures 30 and 31. Also plotted on Figure 30 are the thickener solids loading from the pilot plant thickener presented in Table 23. None of the pilot plant operating periods produced an underflow solids concentration equal to that predicted by the laboratory tests. Highest underflow solids from the pilot plant thickener was 3.0%; on Figure 30, this operating point appears in an area where the bench-scale tests predicted an underflow concentration in excess of 5.0%.

Figure 31 shows the relation between thickener feed solids concentration and the maximum predicted underflow concentration; the trend is substantiated by the pilot thickener results.

Phase 3 (Low Alum; with Polymer)

Results of gravity thickener performance during Phase 3 are summarized in Table 24. As shown in Table 24, two stable operating periods were obtained. Feed solids concentration of 32-37 g/ℓ provided resultant underflow solids concentrations of 44-48 g/ℓ. Solids loading rates of 13.5-21.2 Kg/day-sq m (2.8-4.3 lb/day-sq ft) did not affect the underflow solids of the alum-primary sludge. Solids capture ranged from 95% to 99%. The data tend to show that higher thickener feed solids concentration increases underflow solids concentration.

Figure 32 shows the results of three separate thickener profiles conducted during Phase 3. Inspection of this figure shows a trend not apparent during the first two phases. In this case, it is apparent that the full depth of the thickener was being utilized in thickening of this sludge, and that as the sludge depth increased, the underflow solids increased.

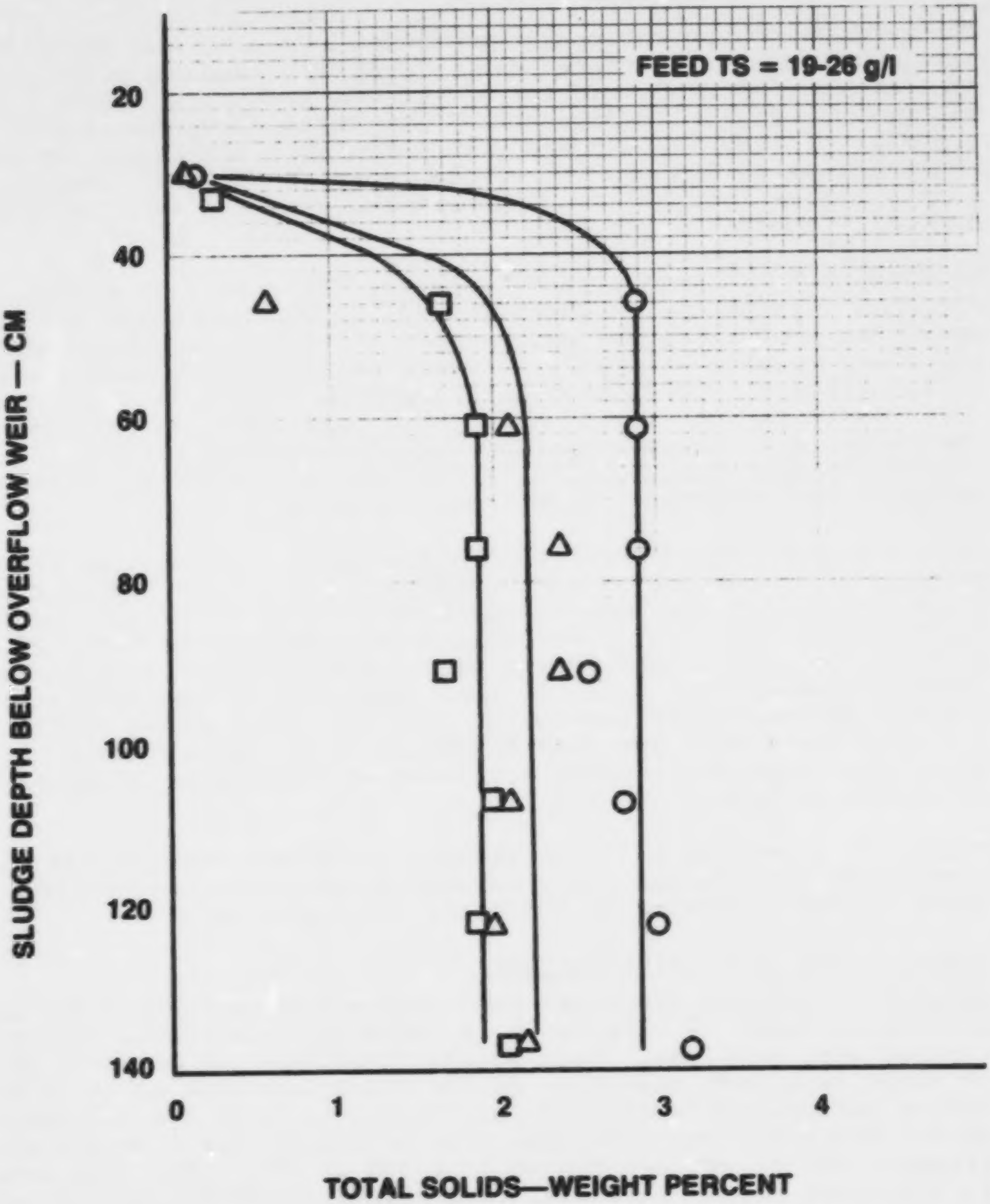

FIGURE 29. ALUM-PRIMARY SLUDGE THICKENER PROFILES PHASE 2.
(Low Alum; with Polymer)

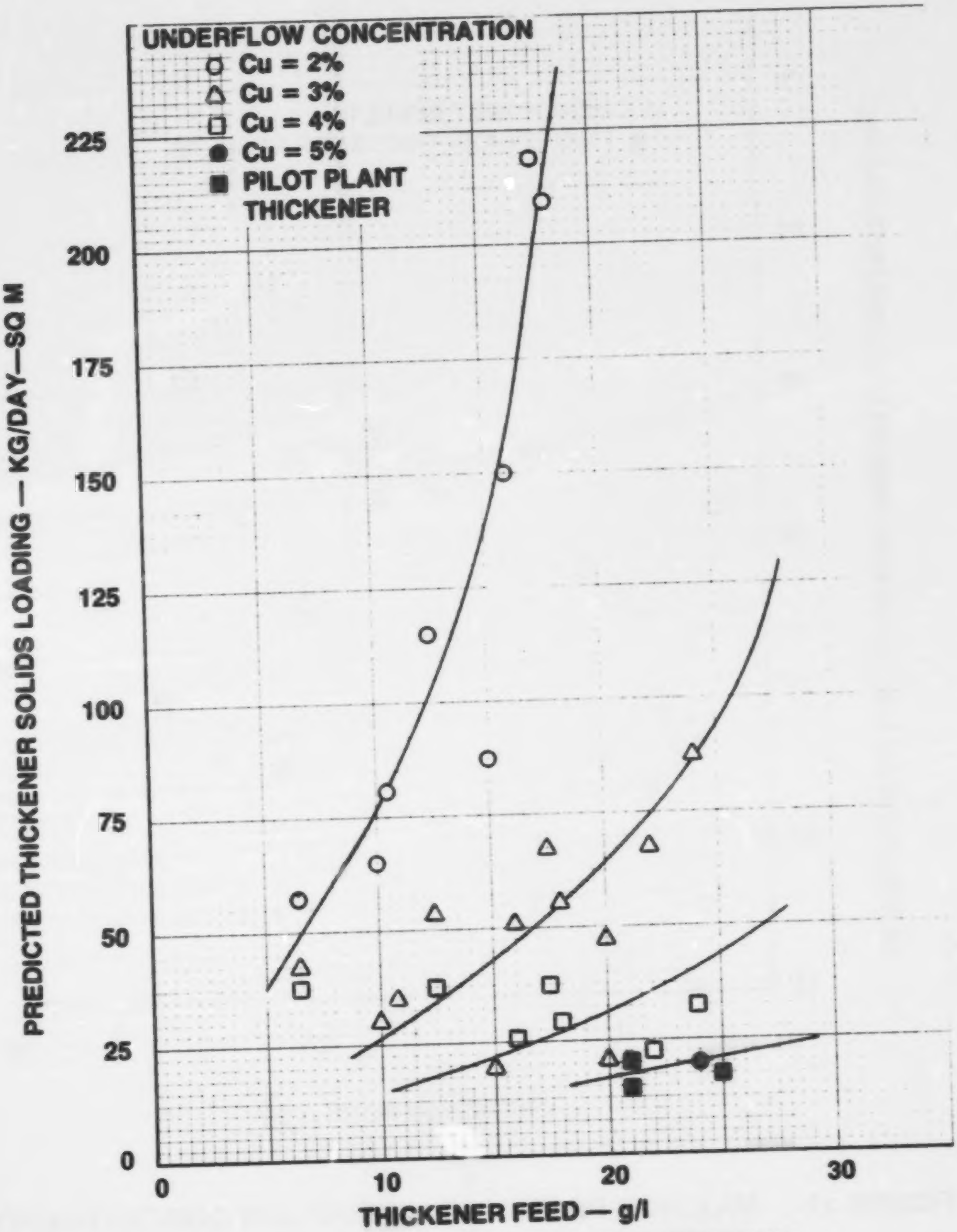

FIGURE 30. PREDICTED TSL VS THICKENER FEED FOR ALUM-PRIMARY SLUDGE — PHASE 2.
(High Alum; with Polymer)

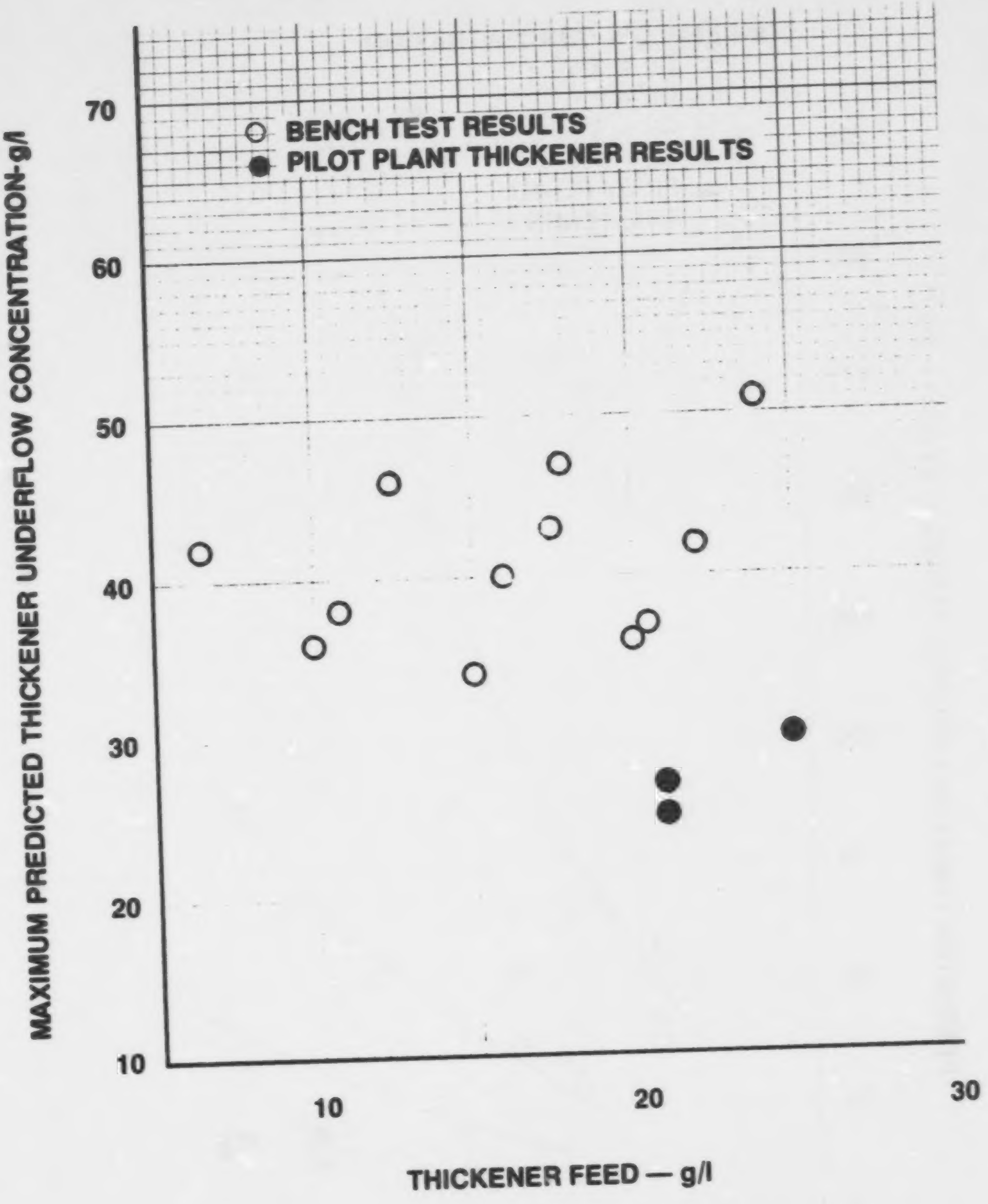

FIGURE 31. MAXIMUM PREDICTED UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION—PHASE 2. (High Alum; with Polymer)

TABLE 24

ALUM-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 3 PILOT PLANT RESULTS
(Low Alum; with Polymer)

Parameter	Values	
Evaluation period, days	5	11
Thickener feed, g/ℓ	32	37
Thickener loading, ℓ/min-sq m	.29	.40
Thickener loading, Kg/day-sq m	13.5	21.2
Thickener sludge age, days	1.6	1.0
Solids capture, percent	.52	.45
Thickener underflow sludge, g/ℓ	95	99
	44	48

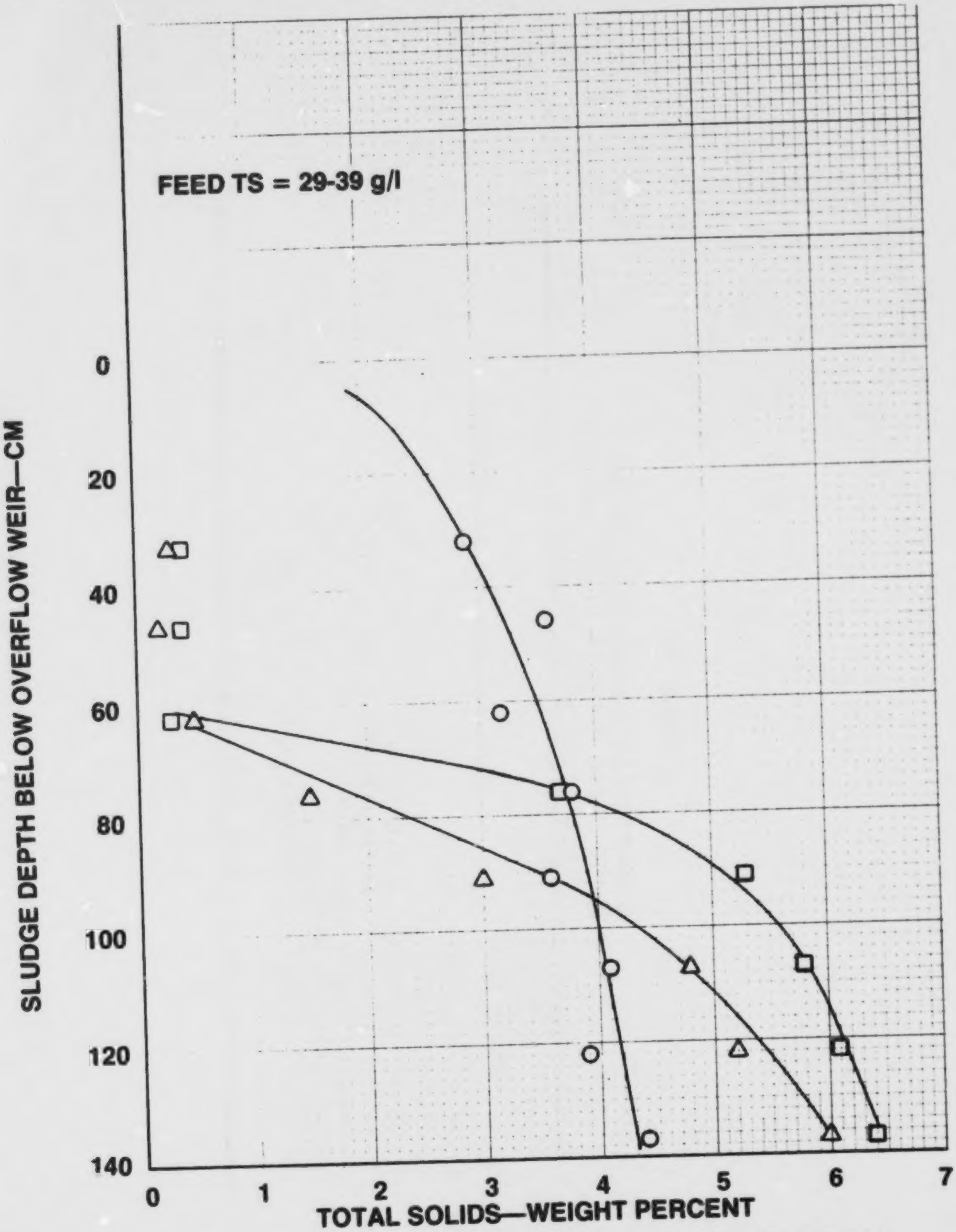

FIGURE 32. ALUM-PRIMARY SLUDGE THICKENER PROFILES PHASE 3.
(Low Alum; with Polymer)

Figures 33 and 34 summarize the results of bench-scale thickening tests conducted on the alum-primary sludge generated during Phase 3. Plotted on Figure 33 are the results of the two stable pilot thickener operating conditions shown in Table 24. Neither of the operating periods produced an underflow comparable to that indicated from the bench-scale tests shown on Figure 33. Figure 34 shows a direct relationship between thickener feed solids concentration and maximum underflow solids concentration. This is substantiated by the results from the pilot plant thickener.

Phase 4 (High Alum; with Polymer)

Results of gravity thickener performance from Phase 4 are summarized in Tables 25 and 26. During Phase 4, thickener performance was closely monitored to determine the effect of feed solids concentration and polymer conditioning of the thickener feed. From Table 25, the effect of diluting the thickener feed is apparent; no improvement in thickener performance was observed. Hydraulic loading was, of course, high when feeding a dilute feed, but solids loadings were generally in line with solids loading experienced when using a more concentrated feed. Solids captures of 98-99% were experienced, even with the highest hydraulic loading of 5.78 ℓpm/sq m (0.14 gal/min-sq ft). Again, the trend of higher underflow concentrations resulting from higher feed solids concentrations was present.

The results of the thickening operations conducted when polymer was added to the thickener feed are summarized in Table 26. The results show that adding polymer to a dilute thickener feed improved thickener performance. Dilution of the feed plus polymer conditioning resulted in an underflow solids concentration of 56 g/ℓ which was the highest achieved for alum-primary type sludge. Future studies of feed dilution and polymer addition should examine the effects of sludge age, sludge depth, thickener feed concentration, polymer type and dosage.

Figure 35 shows the same pattern of gravity thickening that was exhibited during Phases 1 and 2. All the profiles on Figure 35 illustrated the effect of sludge depth on underflow concentrations. Only 30-40 cm (1.0-1.3 ft) sludge depth was required to achieve maximum underflow concentrations; sludge depths greater than 40 cm (1.3 ft) were of little use in increasing the solids concentration.

The results of bench-scale thickening tests conducted on alum-primary sludge generated during Phase 4 are given on Figures 36 and 37. Also plotted on these two figures are the results of pilot plant thickening performance from six stable operating periods. With the exception of the period representing polymer feed to the thickener, none of the pilot plant operating periods

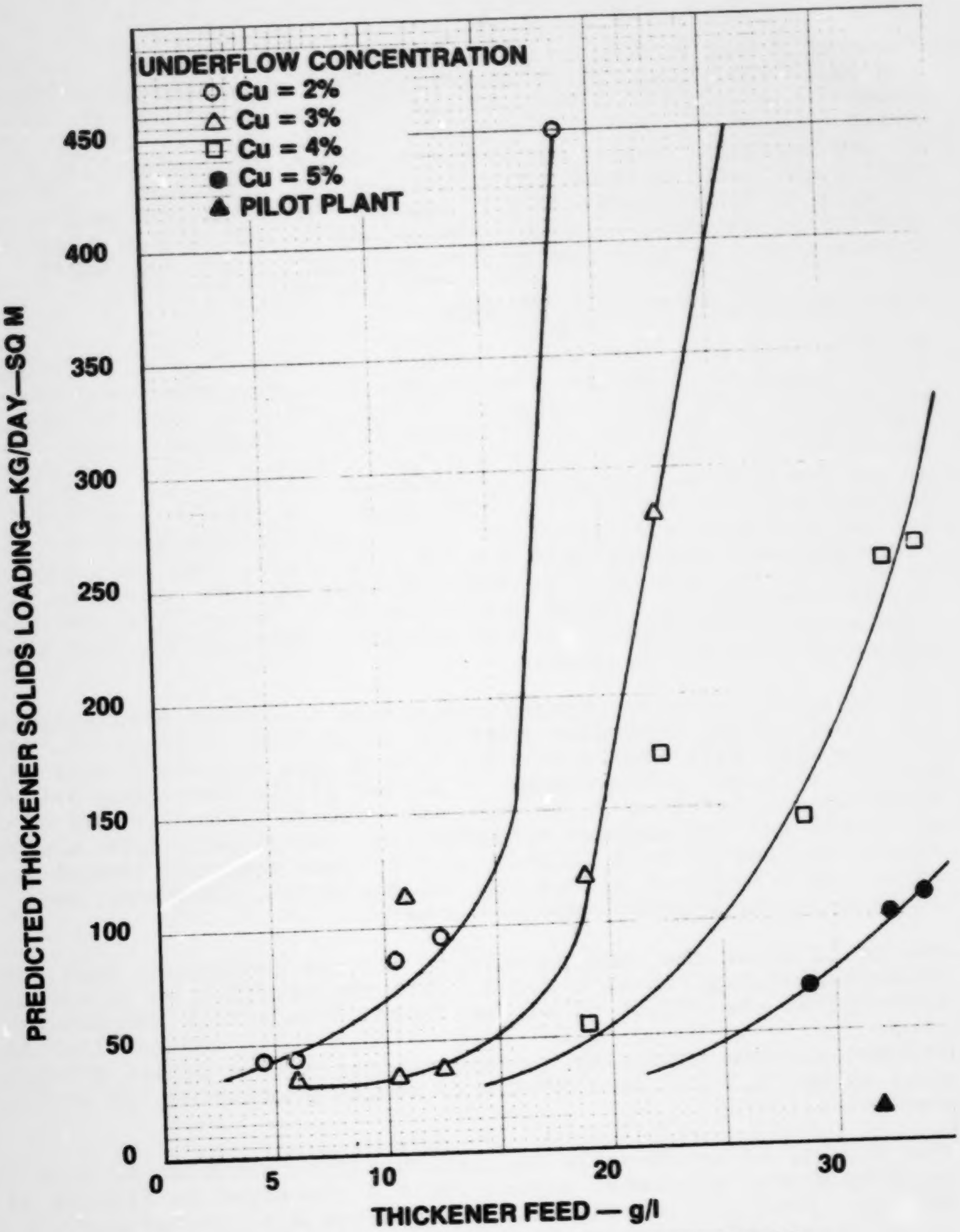

FIGURE 33. PREDICTED TSL VS THICKENER FEED FOR ALUM-PRIMARY SLUDGE—PHASE 3.
(Low Alum; with Polymer)

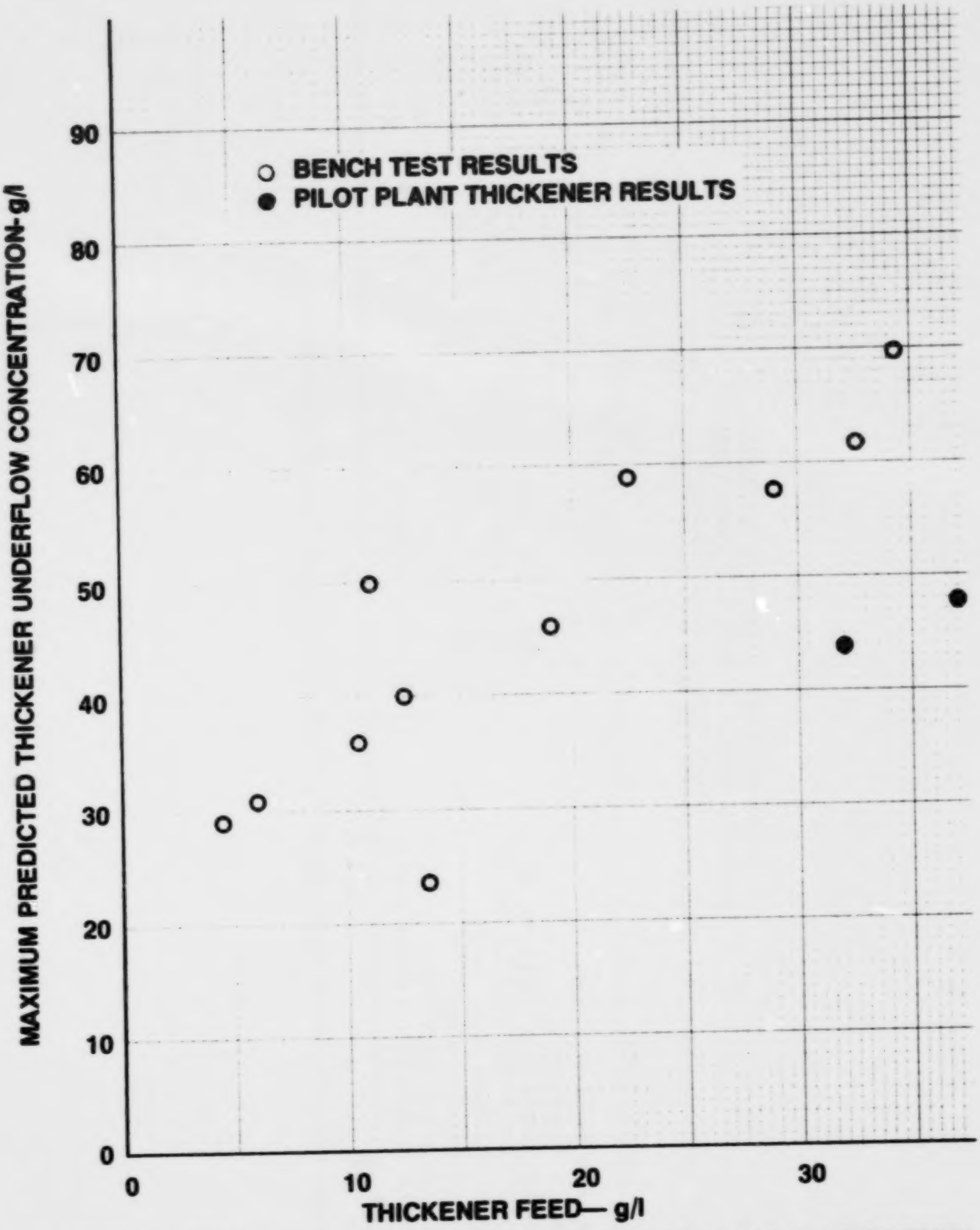

FIGURE 34. MAXIMUM PREDICTED UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION—PHASE 3. (Low Alum; with Polymer)

TABLE 25

ALUM-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 4 PILOT PLANT RESULTS
(High Alum; with Polymer)

Parameter	Values				
Evaluation period, days	5	5	4	2	3
Thickener feed, g/ℓ	36	22	8*	4*	3*
Thickener loading, ℓ/min-sq m	.22	.59	1.95	3.51	5.78
Thickener loading, Kg/day-sq m	11.4	18.7	22.1	22.4	21.2
Thickener sludge age, days	3.42	.07	.68	.55	.27
Thickener sludge depth, m	1.02	.05	.65	.58	.28
Solids capture, percent	99	99	98	98	99
Thickener underflow sludge, g/ℓ	39	29	30	27	26

*Dilution of thickener feed with flocculating clarifier effluent

TABLE 26

ALUM-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 4 PILOT PLANT RESULTS
(High Alum; with Polymer)

Parameter	Values (Polymer Fed to Thickener)
Evaluation period, days	3
Thickener feed, g/ℓ	2*
Thickener loading, ℓ/min-sq m	6.37
Thickener loading, Kg/day-sq m	21.9
Thickener sludge age, days	0.56
Thickener sludge depth, m	0.29
Polymer dosage, % by weight	0.07+
Solids capture, percent	97
Thickener underflow sludge, g/ℓ	56

*Dilution of thickener feed with flocculating clarifier effluent.

+Dow AP-30.

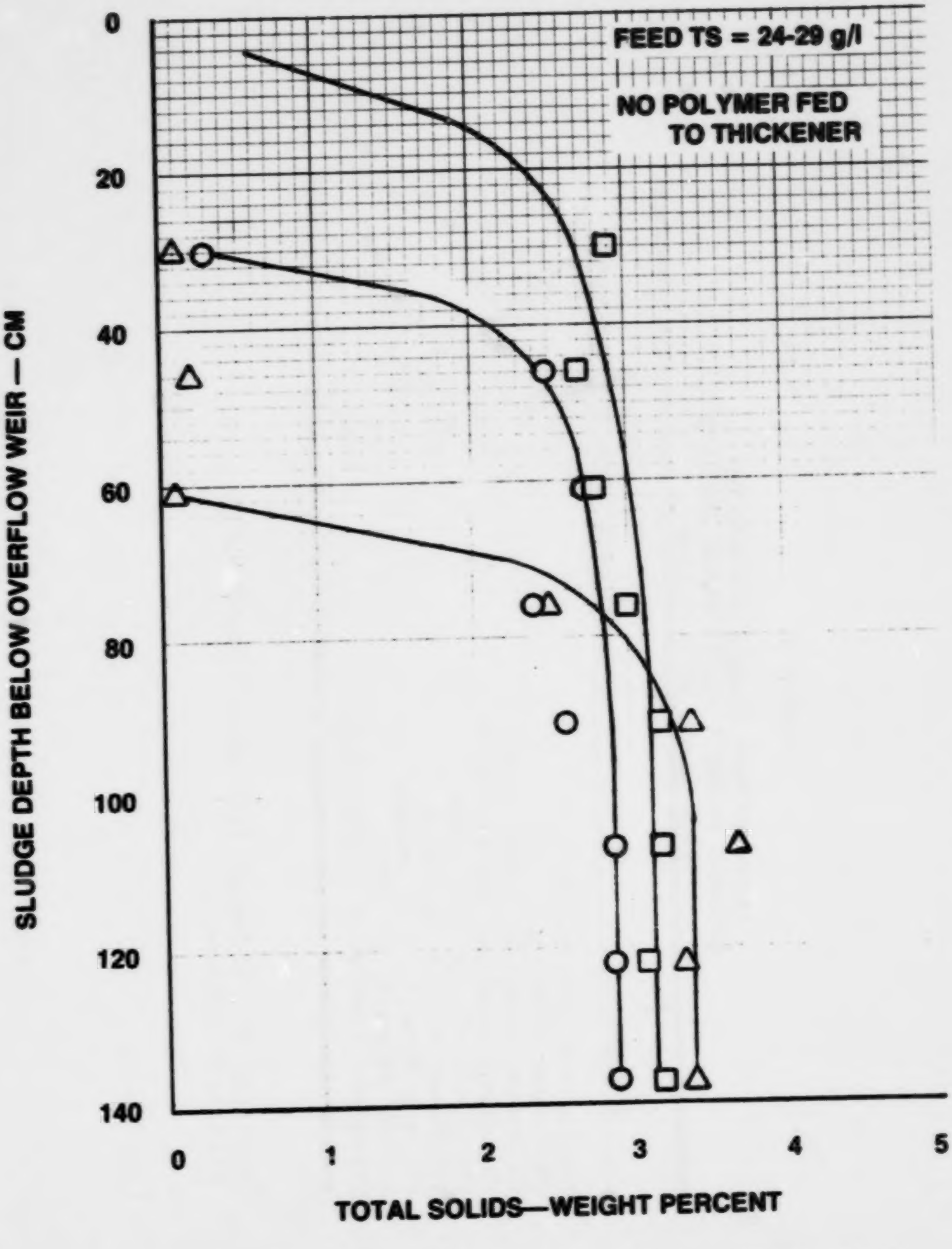

FIGURE 35. ALUM-PRIMARY SLUDGE THICKENER PROFILES PHASE 4.
(High Alum; with Polymer)

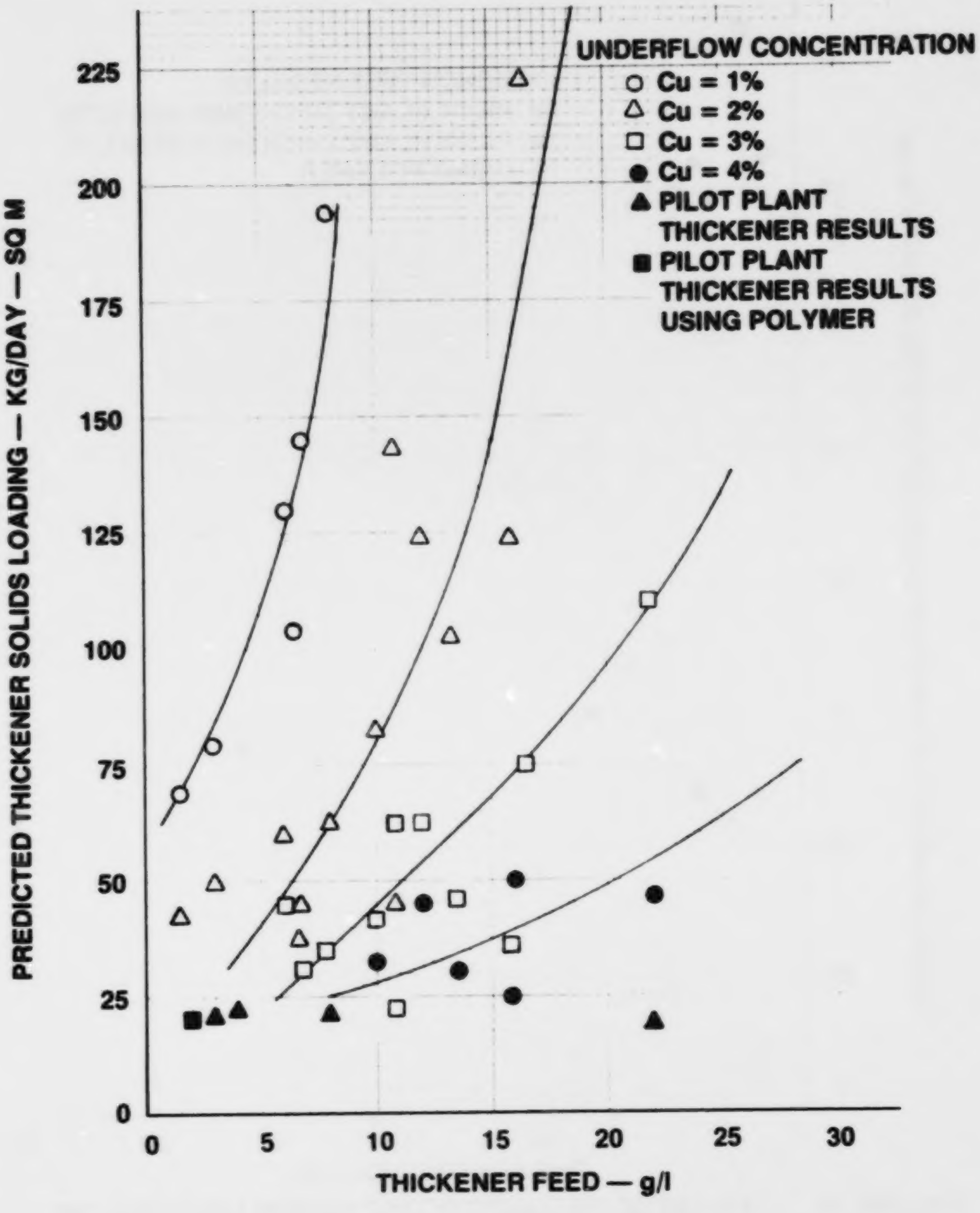

FIGURE 36 PREDICTED TSL VS THICKENER FEED FOR ALUM—PRIMARY SLUDGE — PHASE 4
(High Alum; with Polymer)

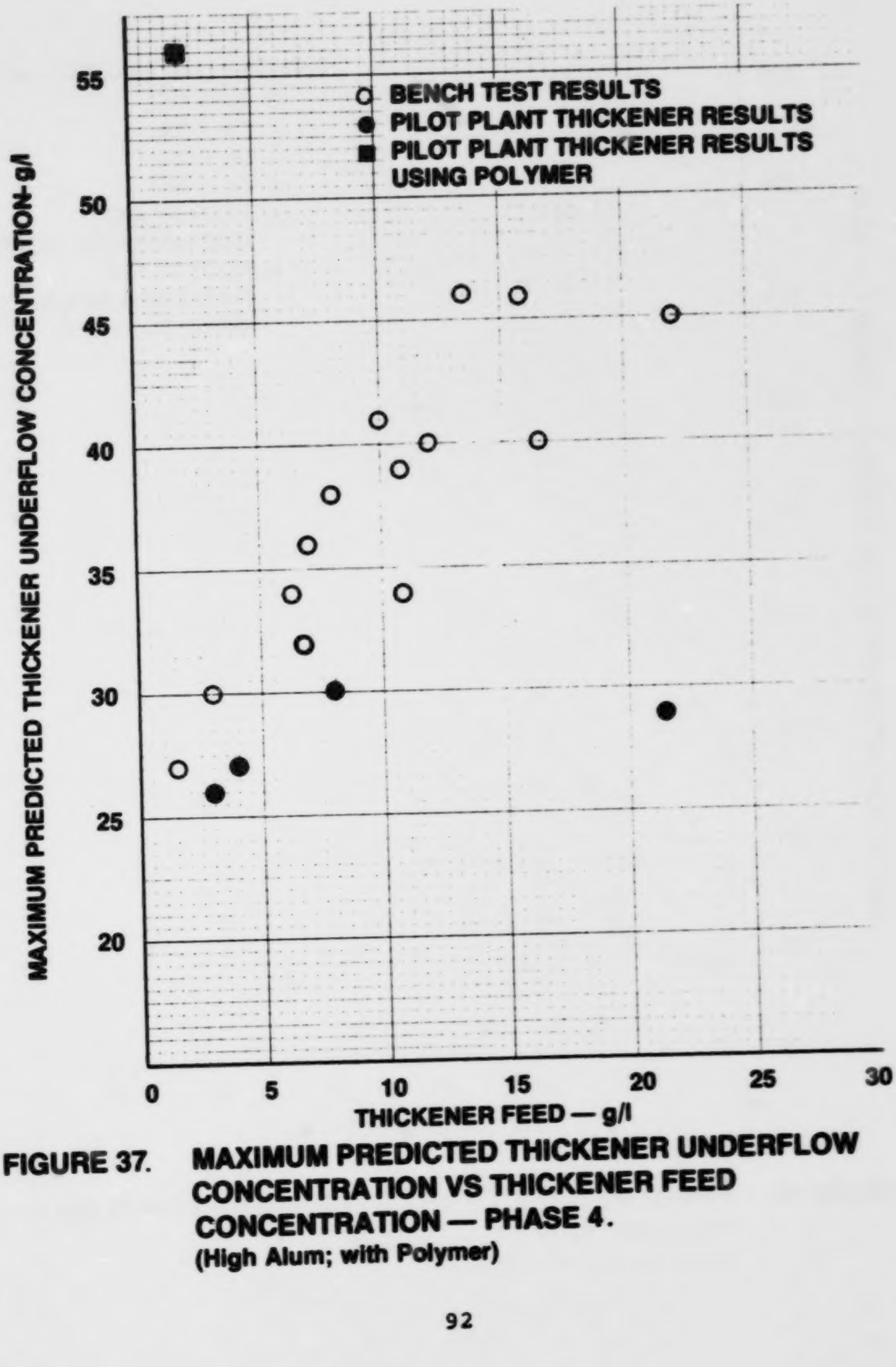

FIGURE 37. MAXIMUM PREDICTED THICKENER UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION — PHASE 4. (High Alum; with Polymer)

produced an underflow solids concentration equal to that predicted by the laboratory tests, although the results of the three periods representing a dilute feed to the thickener approach the laboratory results. Figure 37 shows the relation between thickener feed solids concentration and the maximum predicted underflow solids concentration; the trend is generally substantiated by the pilot plant results.

FERRIC-PRIMARY SLUDGE

Phase 1 (High Ferric; no Polymer)

In Table 27, two stable operating periods from Phase 1 summarize the gravity thickener performance during this phase. The long sludge ages and relatively low solids loading rates for this period were due to operational inexperience.

The first operating period (five days) had an exorbitantly long sludge age; thickener feed solids of 38 g/ℓ resulted in an underflow solids concentration of 48 g/ℓ. For a shorter sludge age and slightly higher thickener solids loading, a feed solids concentration of 18 g/ℓ resulted in an underflow of 64 g/ℓ. However, in this case solids capture was only 87%.

Figure 38 shows the effect of sludge depth in the thickener on underflow solids concentration. Both curves on Figure 38 show the same pattern; the first 40-50 cm (1.3-1.6 ft) of sludge depth produced the majority of the thickening.

Figures 39 and 40 present the results of bench-scale thickening tests performed on ferric-primary sludge generated during Phase 1. Also plotted on Figure 39 are the two pilot plant operating points. Results from one of the thickener operating periods appears consistent with the bench-scale test results, whereas, the pilot performance measured at a long thickener sludge age was significantly poorer than bench-scale results.

Figure 40 illustrates a definite trend in the relationship between thickener feed solids and thickener underflow solids. Pilot plant thickener operation did not substantiate this, but this was probably due to an excessively long thickener sludge age during one of the operating periods.

Phase 2 (High Ferric; with Polymer)

Table 28 summarizes the results of the pilot plant thickening studies conducted on ferric-primary sludge during Phase 2. Four separate operating periods are shown, each of relatively long duration. Thickener feed solids ranged from 17-22 g/ℓ with underflows ranging from 45-57 g/ℓ. Inspection of the three operating periods in which the thickener feed solids were identical (22 g/ℓ) shows an interesting phenomenon. For the

TABLE 27

FERRIC-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 1 PILOT PLANT RESULTS
(High Ferric; no Polymer)

Parameter	Values	
Evaluation period, days	5	8
Thickener feed, g/ℓ	38	18
Thickener loading, ℓ/min-sq m	0.13	0.36
Thickener loading, Kg/day-sq m	7.1	9.2
Thickener sludge age, days	6.5	3.4
Thickener sludge depth, m	1.02	0.62
Solids capture, percent	99	87
Thickener underflow sludge, g/ℓ	48	64

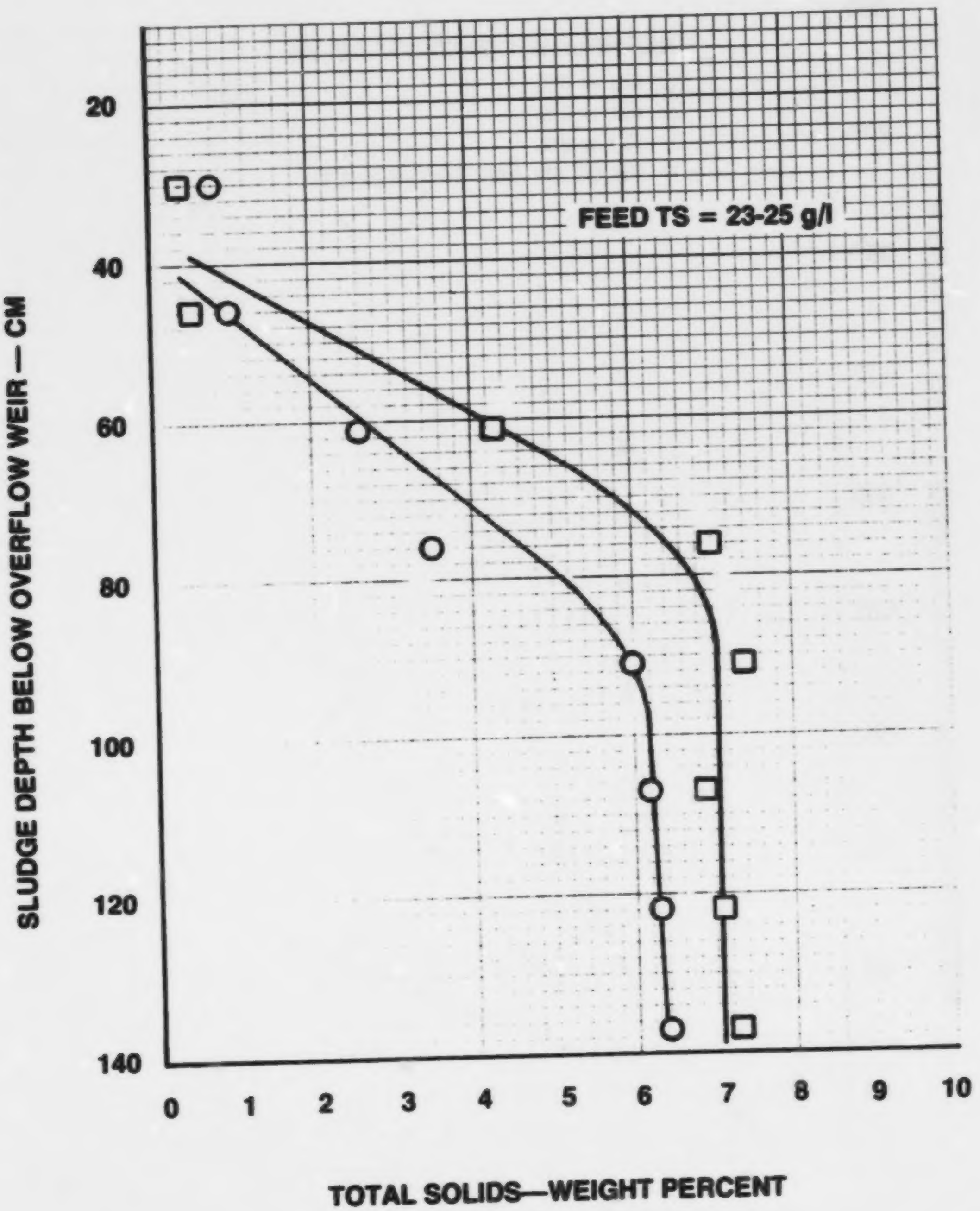

FIGURE 38. FERRIC-PRIMARY SLUDGE THICKENER PROFILES PHASE 1.
(High Ferric; no Polymer)

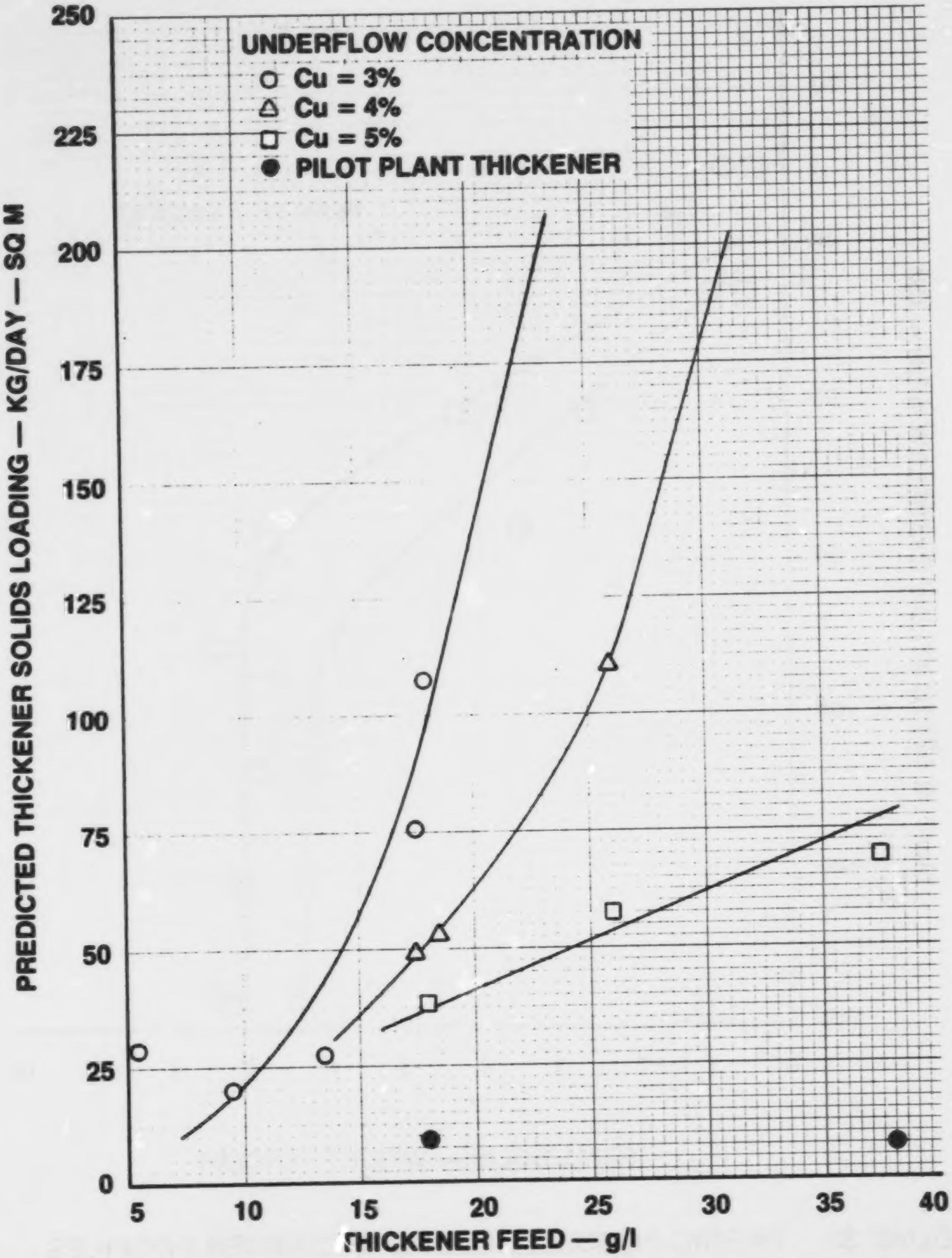

FIGURE 39. PREDICTED TSL VS THICKENER FEED FOR FERRIC — PRIMARY SLUDGE — PHASE 1.
(High Ferric; no Polymer)

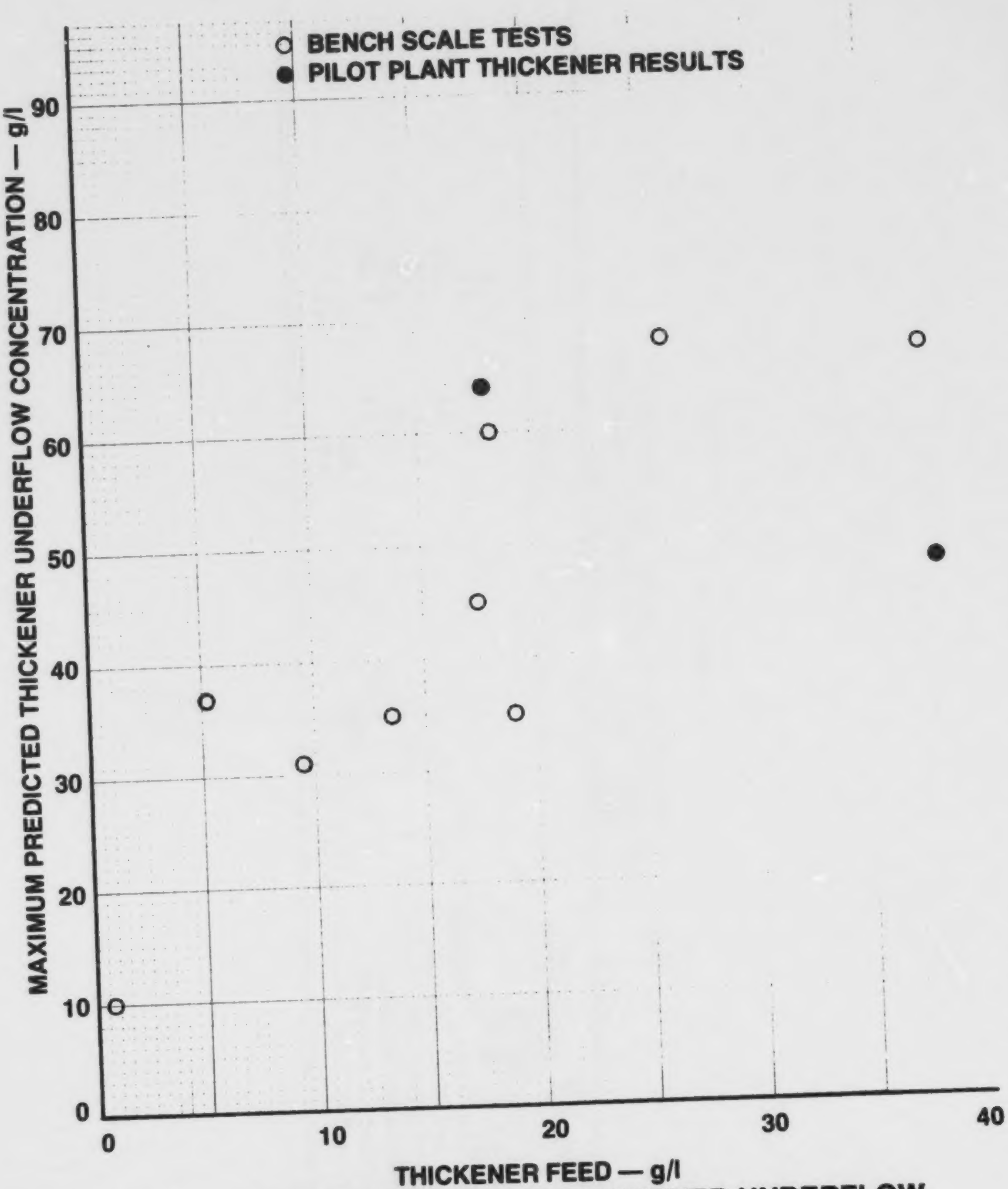

FIGURE 40. MAXIMUM PREDICTED THICKENER UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION — PHASE 1. (High Ferric; no Polymer)

TABLE 28

FERRIC-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 2 PILOT PLANT RESULTS
(High Ferric; with Polymer)

Parameter	Values			
Evaluation period, days	7	5	8	8
Thickener feed, g/ℓ	17	22	22	22
Thickener loading, ℓ/min-sq m	.52	.53	.44	.56
Thickener loading, Kg/day-sq m	12.9	16.5	13.9	17.5
Thickener sludge age, days	3.3	2.5	2.8	2.7
Thickener sludge depth, m	1.14	.93	.83	1.22
Solids capture, percent	99	99	99	97
Thickener underflow sludge, g/ℓ	46	53	57	45

same sludge age (2.5-2.8 days) and the same feed solids, a higher thickener solids loading caused a reduction in the thickener solids underflow concentration. From the data, it can be seen that exceeding a loading of 16.5 Kg/day-sq m (3.4 lb/day sq ft) caused a deterioration in the underflow solids (53 to 45 g/ℓ). For all loadings except the highest one, solids capture was 99%; for the highest loading, solids capture dropped slightly to 97%.

Figure 41 shows three typical thickener solids profiles from the ferric-primary sludge of Phase 2. Clearly, it can be seen that the full depth of the thickener blanket was being utilized in thickening this type of sludge; as the sludge blanket depth increased, the underflow total solids concentration increased.

Figures 42 and 43 summarize the results of the bench-scale thickening tests of ferric-primary sludge from Phase 2. Also shown on Figure 42 are the results of the four pilot plant thickening periods presented in Table 28. Generally, pilot plant results were consistent with predicted performance from bench-scale tests. Figure 43 shows a relationship between the thickener feed solids and predicted underflow solids. This is basically substantiated by the pilot plant thickener results presented in Table 28.

Phase 3 (Low Ferric; with Polymer)

Average pilot plant thickening results for three operating periods from Phase 3 are presented in Table 29. Thickener feed solids ranged from 33-52 g/ℓ, with underflow solids ranging from 43-62 g/ℓ. Higher thickener solids loadings were achieved during this phase, and underflows produced were generally slightly higher than those produced in the prior two phases. High solids captures were also experienced (98-99%).

Figure 44 shows three typical thickener solids profiles. Near maximum underflow concentration was achieved in the first 40 cm (1.3 ft) of blanket depth with only nominal increases in underflow concentration occurring with further increases in blanket depth.

Bench-scale thickening results are summarized on Figures 45 and 46. Shown on Figure 45 are the pilot plant thickener results. Pilot thickener results are generally inconsistent with predicted performance from bench-scale results. Figure 46 shows a definite relationship between the thickener feed and predicted thickener underflow concentration. This trend was substantiated by the results of the pilot plant thickener.

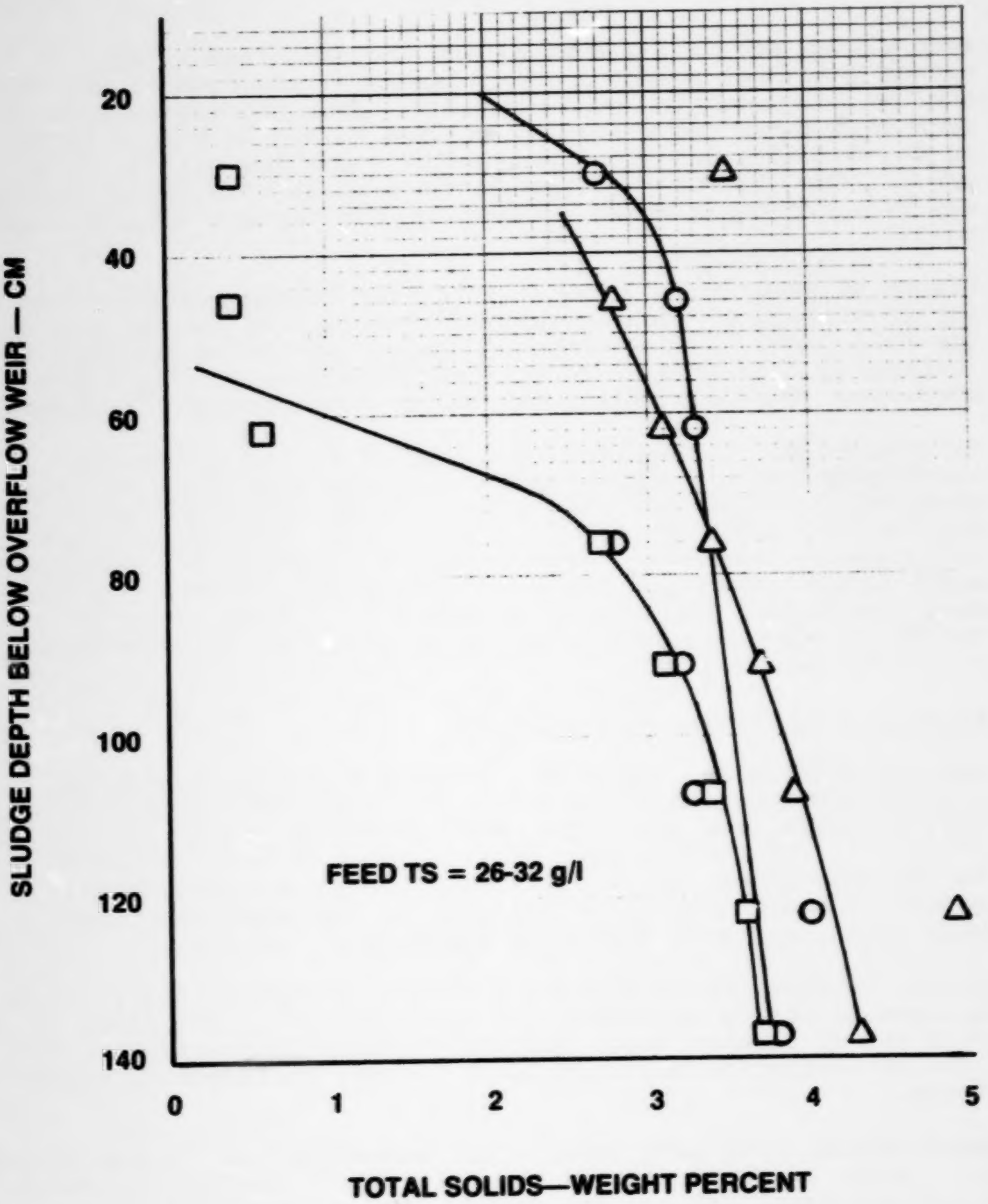

FIGURE 41. FERRIC-PRIMARY SLUDGE THICKENER PROFILES PHASE 2.
(High Ferric; with Polymer)

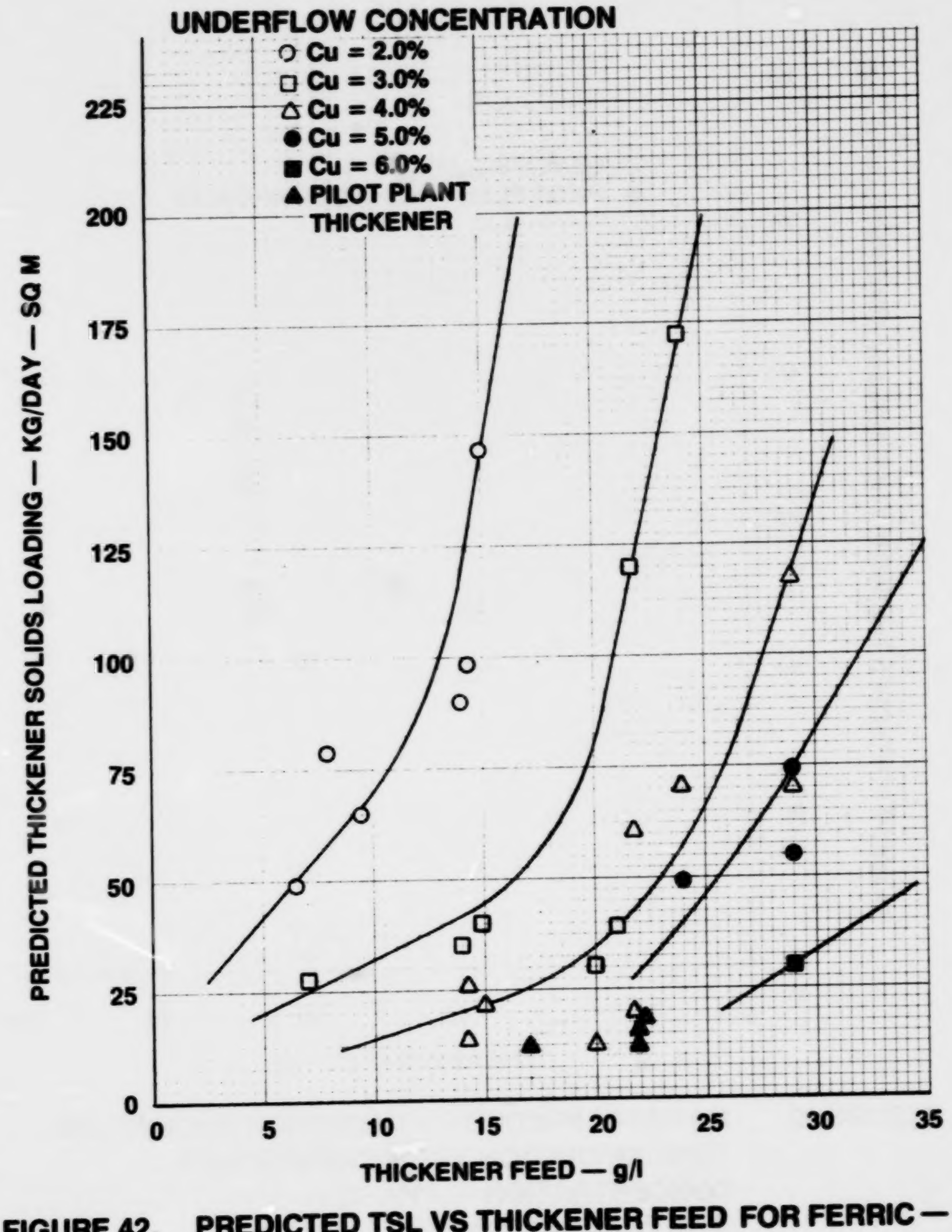

FIGURE 42. PREDICTED TSL VS THICKENER FEED FOR FERRIC — PRIMARY SLUDGE — PHASE 2.
(High Ferric; with Polymer)

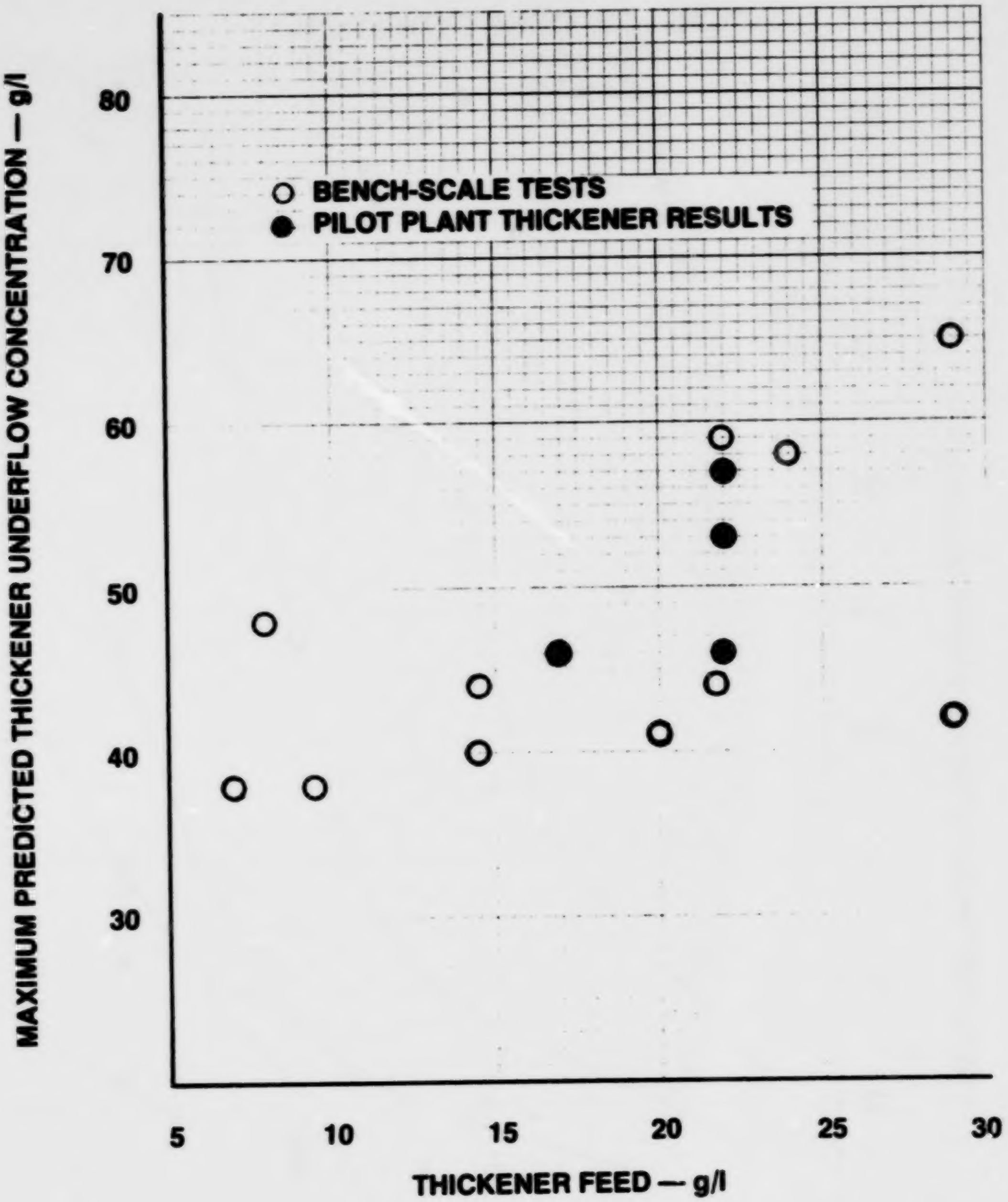

FIGURE 43. MAXIMUM PREDICTED THICKENER UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION — PHASE 2.
(High Ferric; with Polymer)

TABLE 29

FERRIC-PRIMARY SLUDGE THICKENING
AVERAGE PHASE 3 PILOT PLANT RESULTS
(Low Ferric; with Polymer)

Parameter	Values		
Evaluation period, days	4	4	8
Thickener feed, g/ℓ	33	52	38
Thickener loading, ℓ/min-sq m	.65	.35	.40
Thickener loading, Kg/day-sq m	30.8	26.1	22.0
Thickener sludge age, days	.35	1.6	1.5
Thickener sludge depth, m	.27	.71	.63
Solids capture, percent	99	99	98
Thickener underflow sludge, g/ℓ	43	62	60

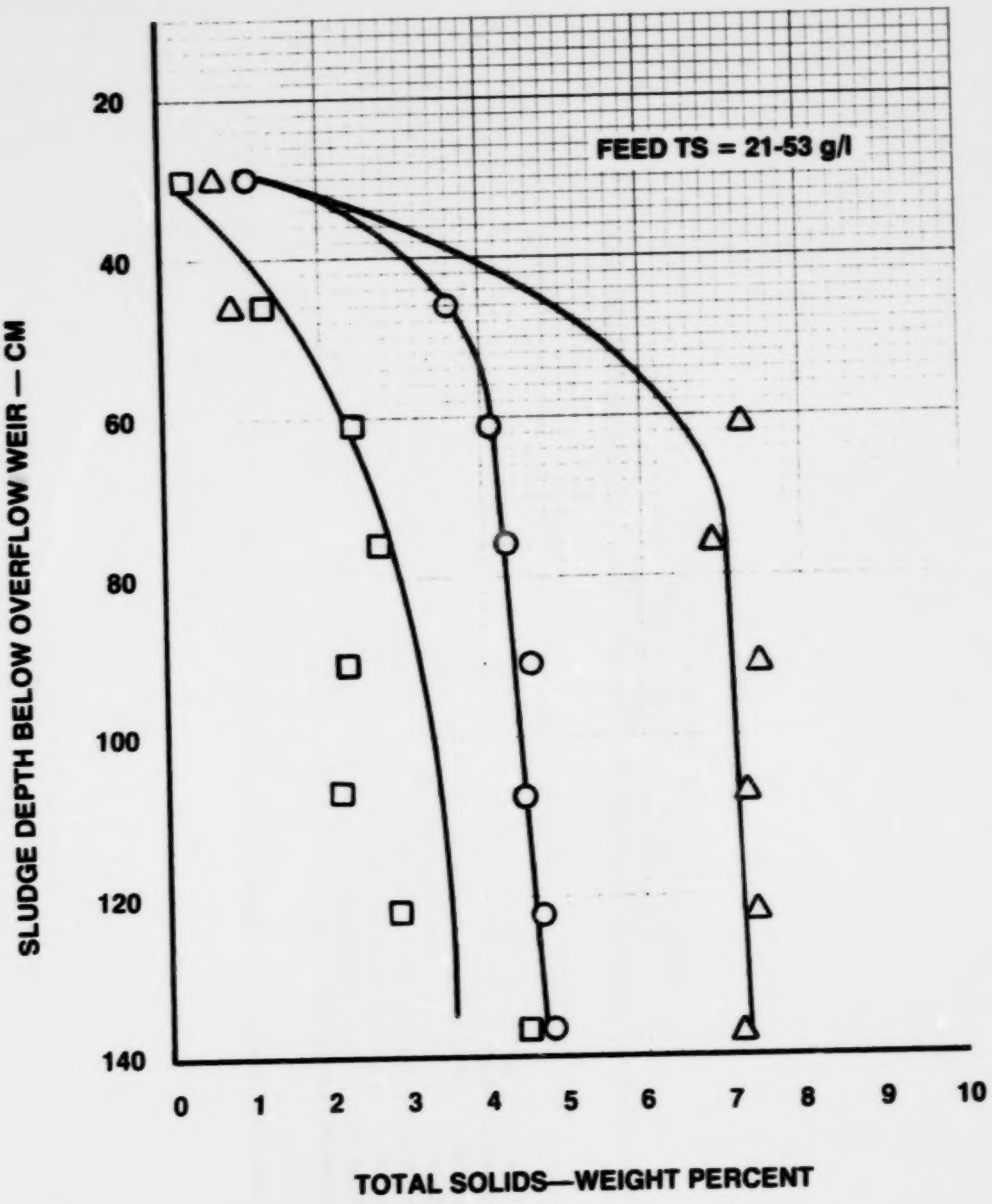

FIGURE 44. FERRIC-PRIMARY SLUDGE THICKENER PROFILES PHASE 3.
(Low Ferric; with Polymer)

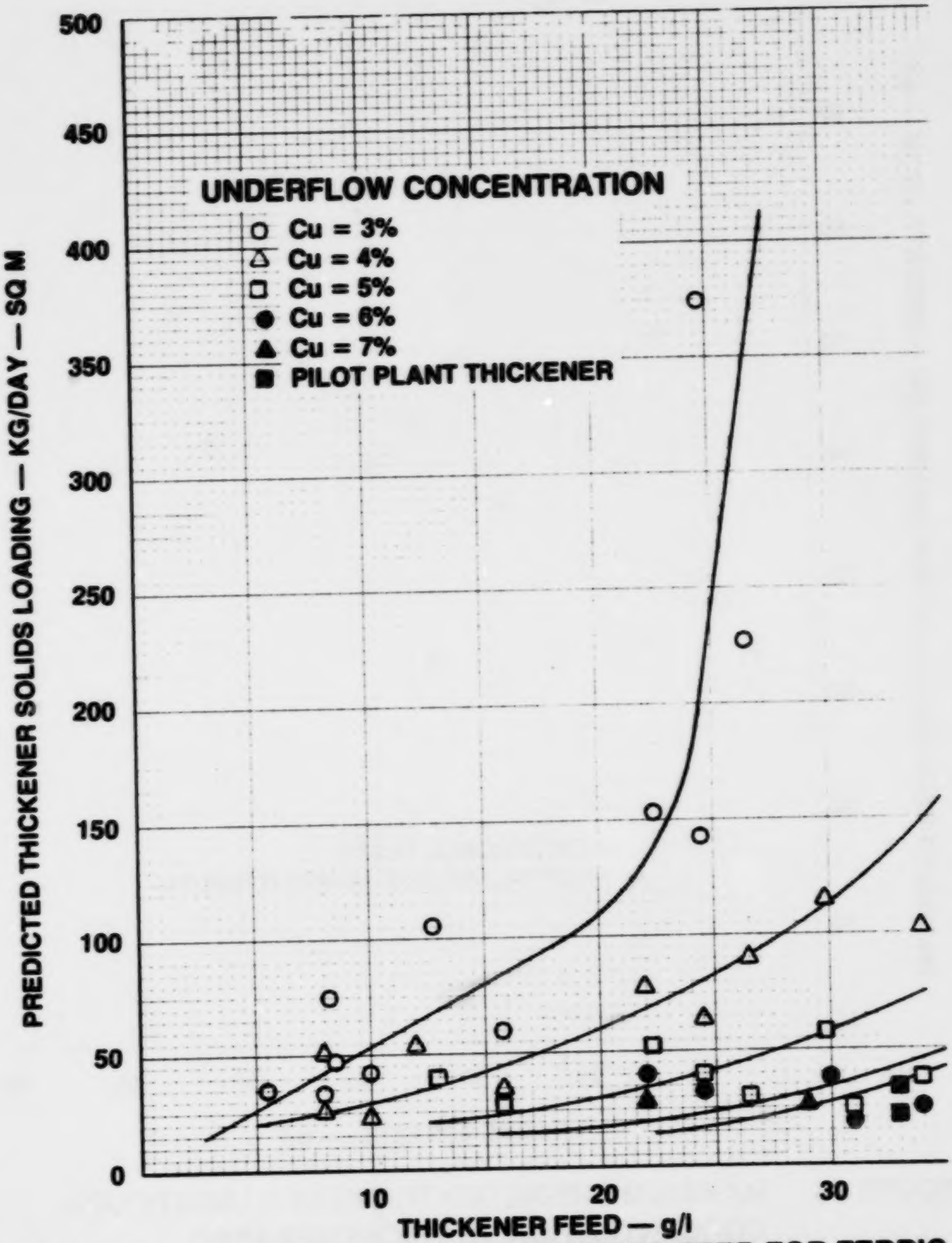

FIGURE 45. PREDICTED TSL VS THICKENER FEED FOR FERRIC — PRIMARY SLUDGE — PHASE 3.
(Low Ferric; with Polymer)

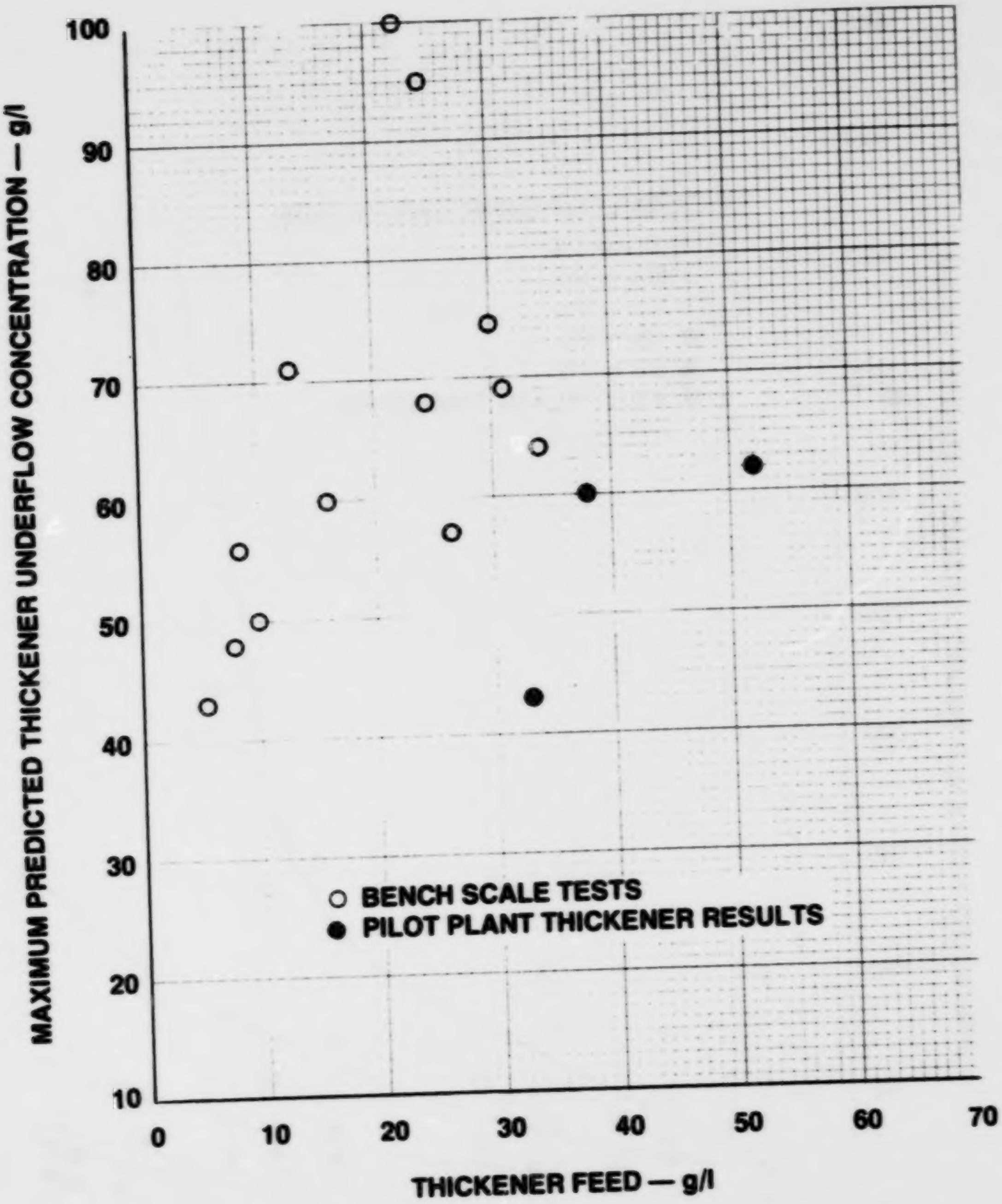

FIGURE 46. MAXIMUM PREDICTED THICKENER UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION — PHASE 3.
(Low Ferric; with Polymer)

PRIMARY SLUDGE

Phase 4

During Phase 4, ferric-primary sludge was no longer produced. Primary sludge instead was produced to attempt to expand the thickening data base.

Table 30 summarizes the results of the three operating periods from Phase 4. As expected, and as indicated by the first two operating periods shown in Table 30, once chemical coagulant addition to the flocculating clarifier was terminated, a sludge with greatly enhanced thickening properties was produced. The first operating period, for example, resulted in an average thickener underflow concentration of 83 g/ℓ.

Interconnecting piping and tank elevations were not designed to permit primary sludge, at such concentrations, to gravity flow from the thickener to the underflow sludge storage tank. Therefore, during the second and third operating periods, thickener solids loading was increased and blanket depth was decreased in an attempt to reduce the thickener underflow concentration.

Figure 47 shows the results of one thickener solids profile conducted on primary sludge. This profile was taken when the sludge level was very low, however, considerable thickening still occurred.

Figures 48 and 49 summarize bench-scale thickening results for primary sludge generated during Phase 4. The results of the three pilot plant operating periods are also plotted on Figure 48. Ignoring the operating period with the underflow of 18 g/ℓ, due to the operating conditions it was run under, pilot plant thickener performance correlated well with predicted bench-scale performance.

Figure 49 shows the results of the thickener feed versus thickener underflow solids concentration. No pattern is discernable.

THICKENING COMPARISONS:ALUM-PRIMARY AND FERRIC-PRIMARY SLUDGES

Figures 50 and 51 show the results of all stable pilot plant thickening periods experienced during the study. Maximum thickener underflow solids are the maximum solids obtained in the thickener underflow for any stable thickener operating period.

Figure 50 shows the impact of percent chemical solids present in the thickener feed sludge on the maximum attainable thickener underflow solids. Because of thickener operational problems, discussed previously, the underflow concentration for 0% chemical solids (i.e., primary sludge) is considered to be lower than could have been achieved.

TABLE 30

PRIMARY SLUDGE THICKENING
AVERAGE PHASE 4 PILOT PLANT RESULTS

Parameter	Values		
Evaluation period, days	17	6	14
Thickener feed, g/ℓ	10	26	10
Thickener loading, ℓ/min-sq m	.52	.67	.97
Thickener loading, Kg/day-sq m	7.3	25.3	13.9
Thickener sludge age, days	2.0	1.0	.02
Thickener sludge depth, m	.24	.49	.02
Solids capture, percent	84	96	87
Thickener underflow sludge, g/ℓ	83	66	18

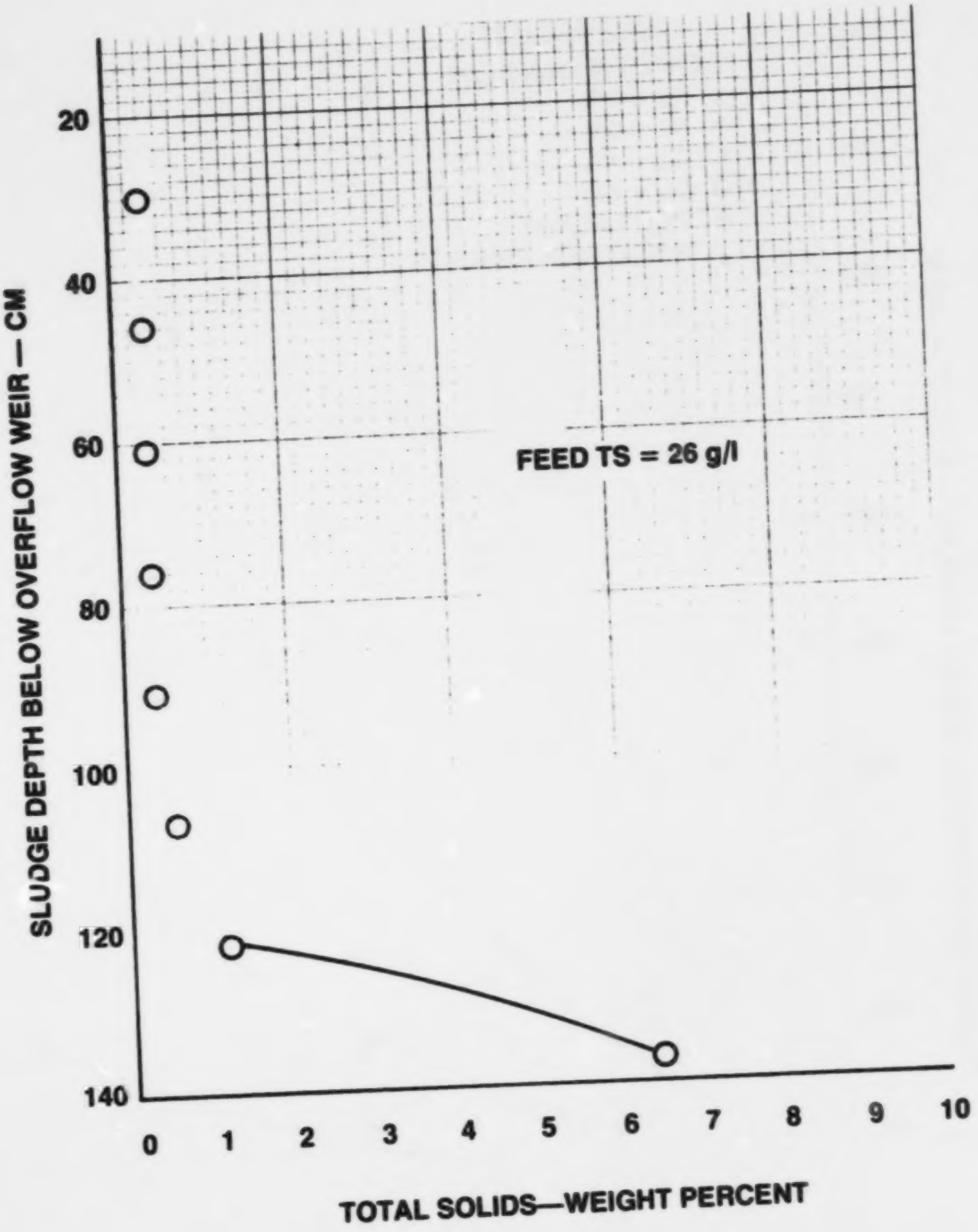

FIGURE 47. PRIMARY SLUDGE THICKENER PROFILES — PHASE 4.

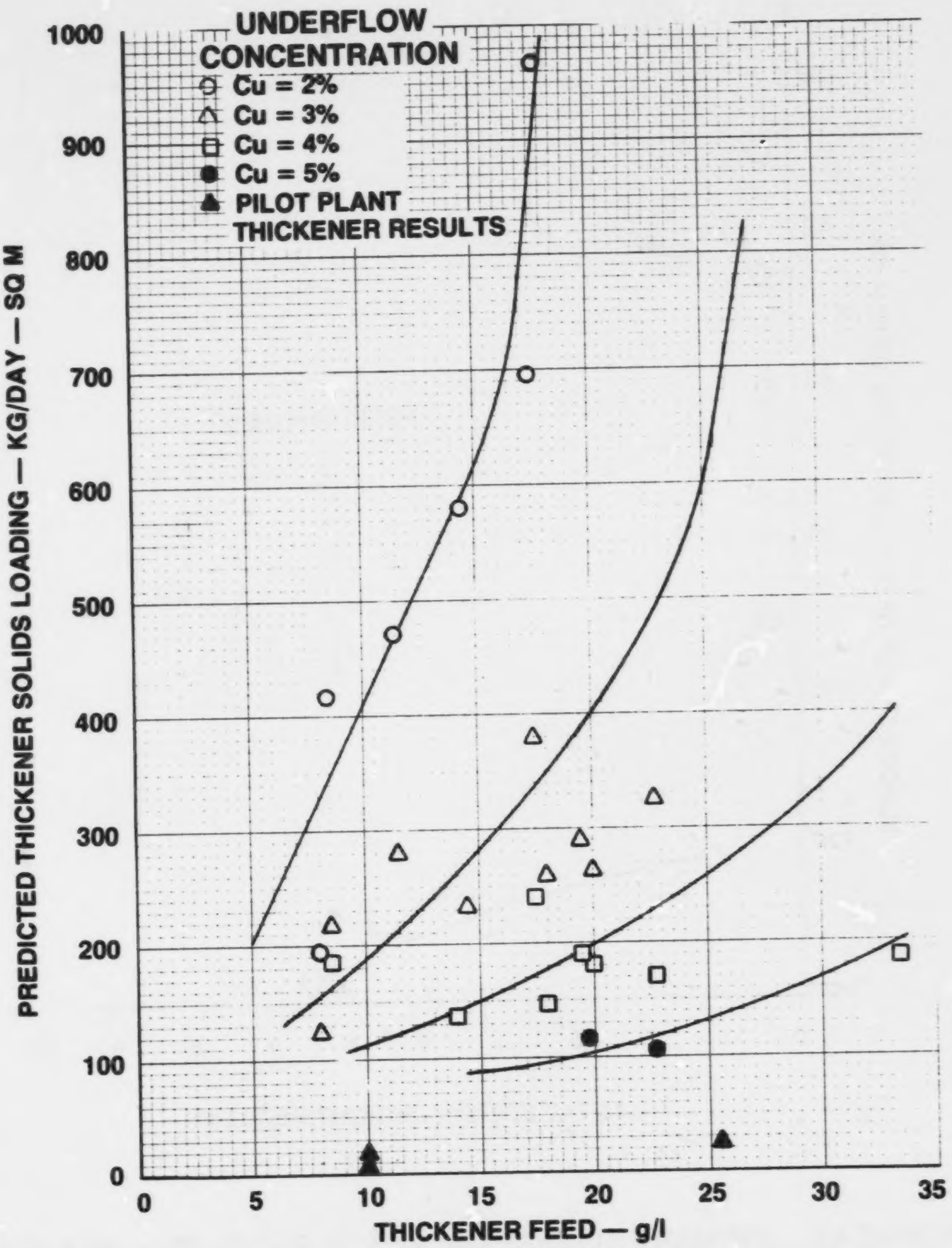

FIGURE 48· PREDICTED TSL VS THICKENER FEED FOR PRIMARY SLUDGE — PHASE 4.

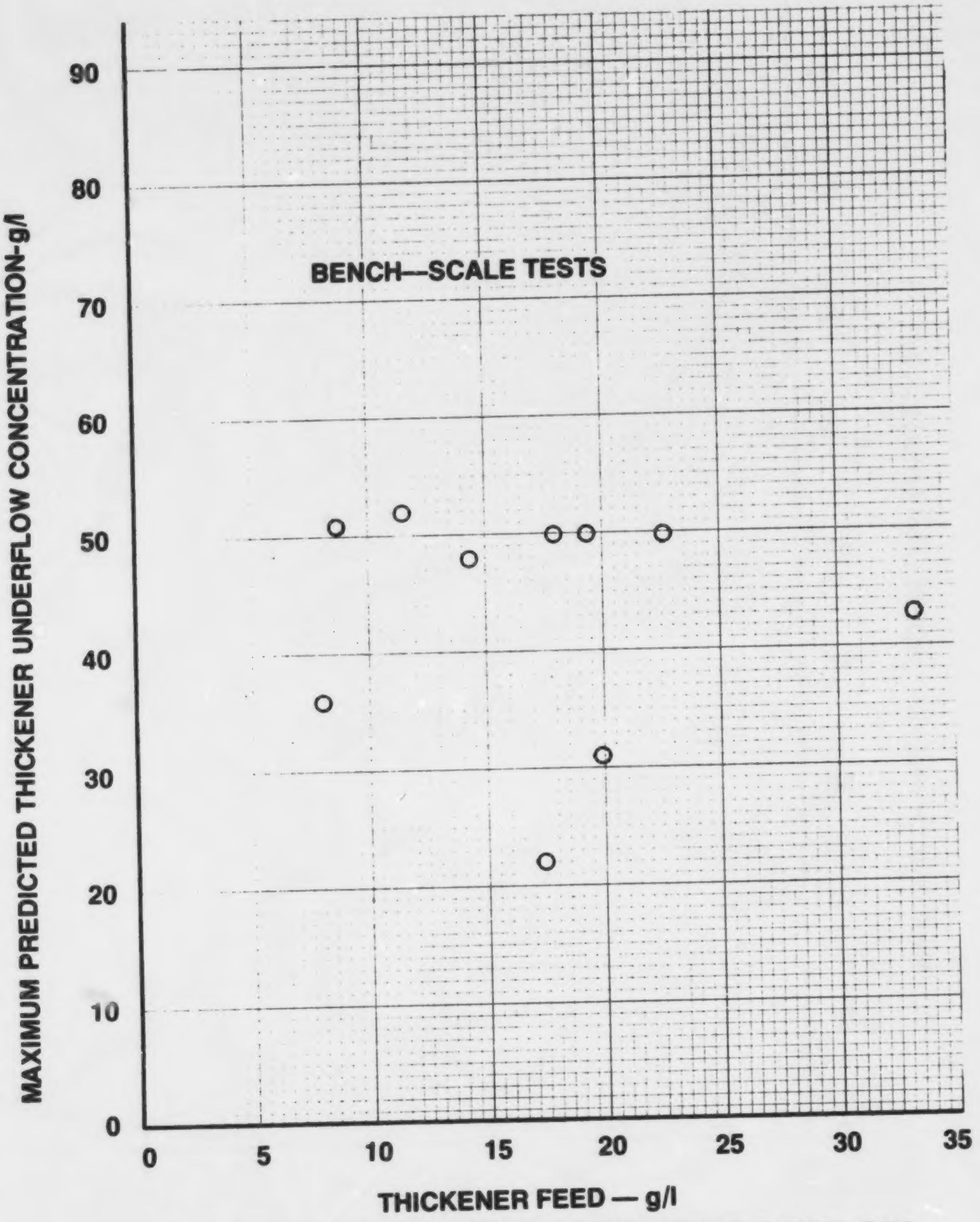

FIGURE 49. MAXIMUM PREDICTED THICKENER UNDERFLOW CONCENTRATION VS THICKENER FEED CONCENTRATION FOR PRIMARY SLUDGE — PHASE 4.

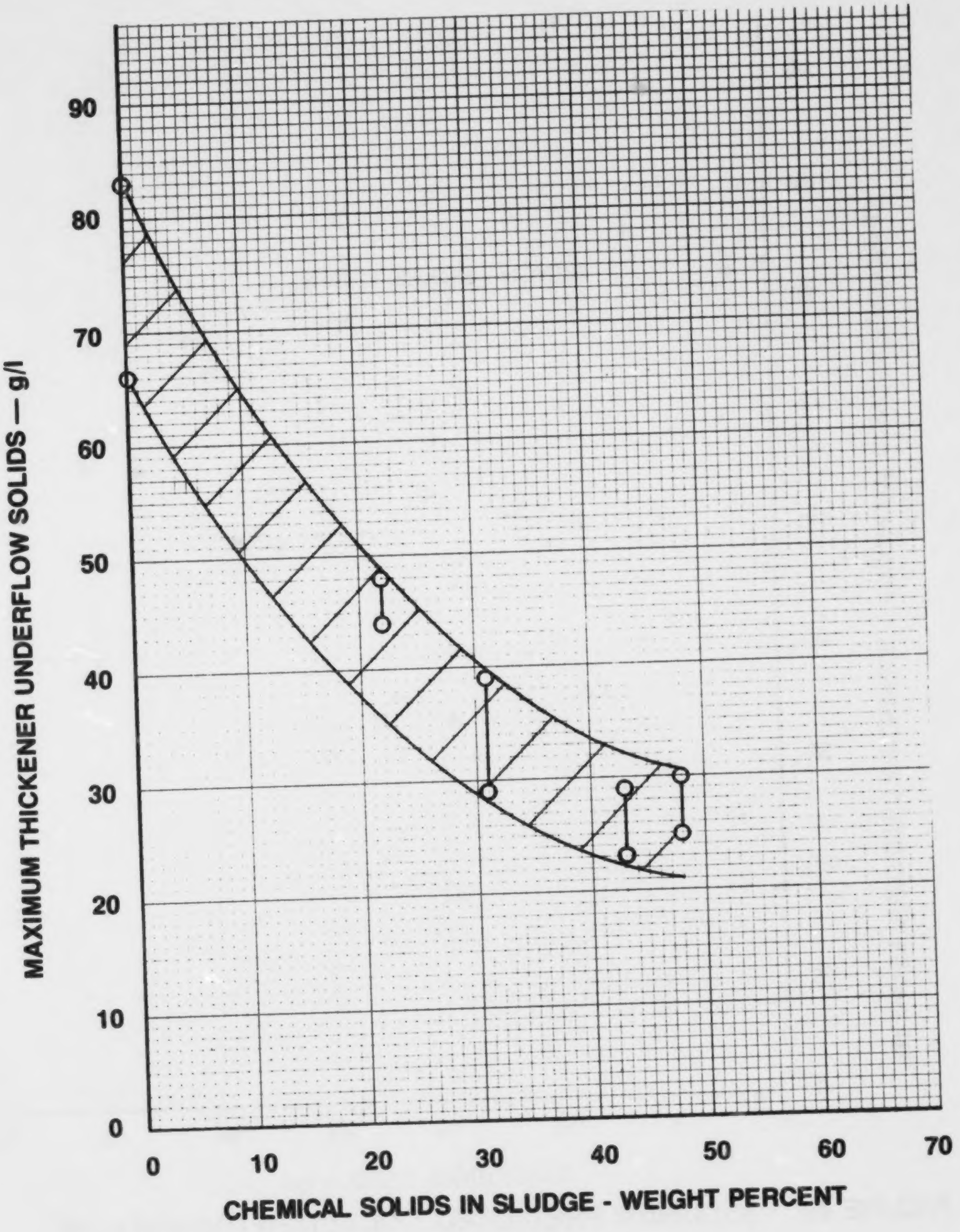

FIGURE 50. RANGE OF THICKENER OPERATING PERIODS FOR ALUM-PRIMARY SLUDGE.

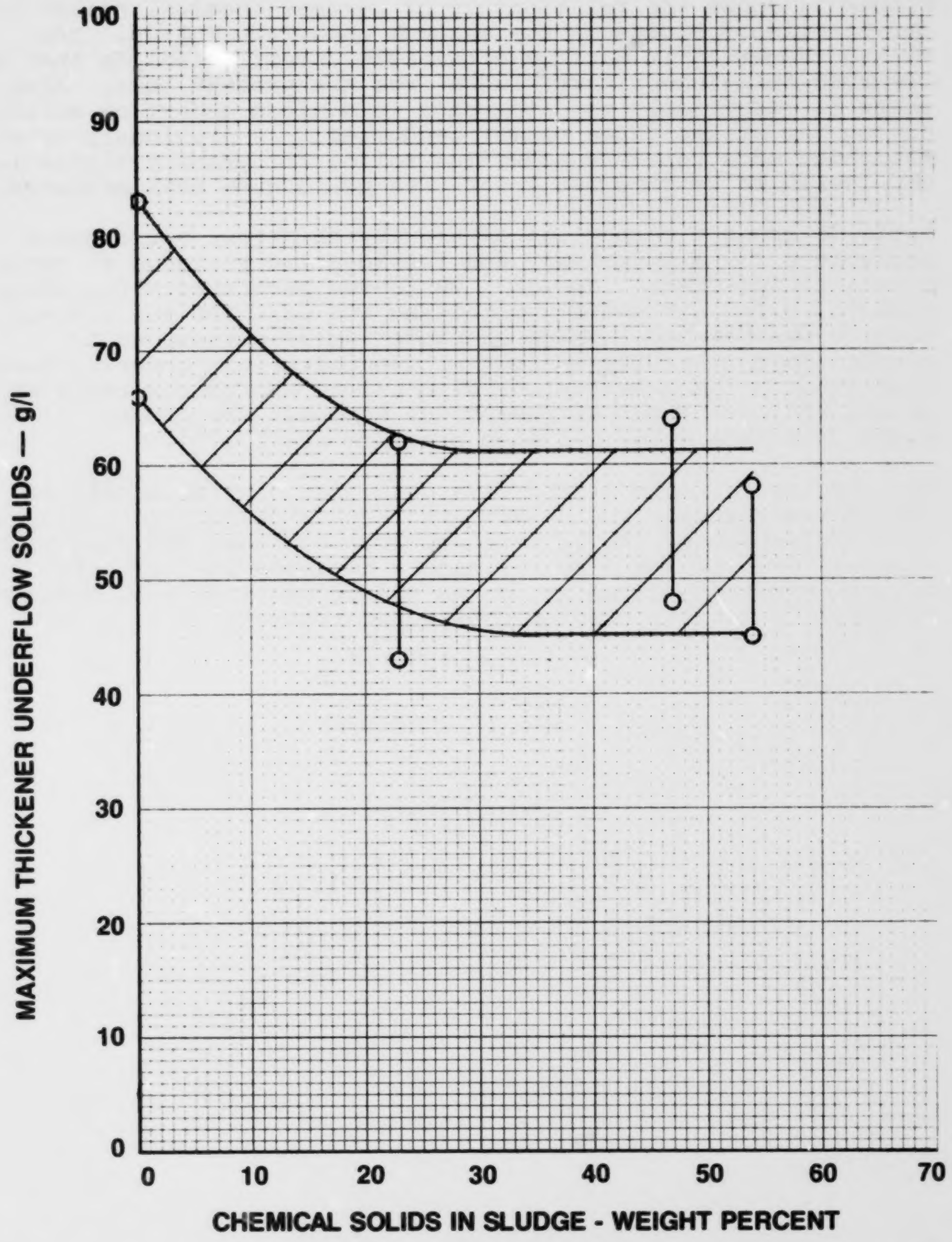

FIGURE 51. RANGE OF THICKENER OPERATING PERIODS FOR FERRIC-PRIMARY SLUDGE.

Figure 51 shows the relationship of percent chemical solids in ferric-primary sludge on the maximum underflow solids. The marked reduction in maximum attainable underflow solids that was observed for alum-primary sludge was not present here. Also, there is essentially no difference in maximum underflow solids during any of the three phases conducted. As previously noted, the range of attainable underflow solids for primary sludge is considered to be somewhat lower than could have been achieved.

Table 31 shows a comparison of gravity thickener performance parameters for alum-primary and ferric-primary sludge at various levels of phosphorus removal. To review performance for these sludges, a raw wastewater containing 100 mg/ℓ TSS and 5.0 mg/ℓ total P was assumed. These characteristics correspond to average Salt Lake City wastewater concentrations observed during this study. The molar ratios of metal to phosphorus required to achieve each level of phosphorus removal were determined from this study and from previous studies[1,2].

The results in Table 31 indicate that, for similar levels of phosphorus removal, the gravity thickening performance parameters for ferric-primary sludge were superior to those for alum-primary sludge. The results also indicate a decrease in thickener underflow solids concentration as additional inorganic coagulant is added to achieve greater phosphorus removal.

TABLE 31

EFFECT OF PHOSPHORUS REMOVAL ON GRAVITY THICKENING PROPERTIES OF ALUM-PRIMARY AND FERRIC-PRIMARY SLUDGE

	ALUM-PRIMARY SLUDGE		FERRIC-PRIMARY SLUDGE	
Total Phosphorus Removal-Percent	Chemical Sludge Weight Percent	Underflow Percent TS	Chemical Sludge Weight Percent	Underflow Percent TS
80	18	4.7	22	5.5
90	23	4.1	28	5.4
95	32	3.3	38	5.3

BASIS: Raw wastewater with 100 mg/ℓ TSS and total phosphorus of 5 mg/ℓ P.
Solids loading rate ~20 Kg/day-sq m

VACUUM FILTRATION DEWATERING

Alum-Primary Sludge

Phase 1 (High Alum; no Polymer)

Alum-primary sludge vacuum filtration data obtained during Phase 1 is given in Table 32.

Form filtration rates are shown on Figure 52. These rates were calculated exclusive of chemical conditioner in the filter cake in order to simplify an evaluation of the effects of conditioning dose. Solid lines through data points were extended over a range of form time which resulted in filter cake dischargeability with a rating of at least "fair". Dotted lines through data points represent the range of form times which resulted in a cake discharge rating of poor to non-dischargeable.

From Figure 52, it can be observed that filtration performance was poor when the alum-primary feed sludge concentration was approximately two weight percent. A lime conditioning dose of 37 weight percent resulted in poor discharge and low filtration rates. One test conducted using a lime dose of 55 weight percent did demonstrate dischargeable cakes; however, filtration rate was also low.

When operating with an alum-primary sludge of approximately three weight percent, dischargeable cakes were produced at a lime dose of 37 weight percent. When the lime dose was reduced to 29 weight percent, a "poor" cake discharge resulted, in addition to lower filtration rates.

Figures 53 and 54 represent full-scale vacuum filter operating curves predicted from the data in Table 32 and on Figure 52. The filtration rates indicated on Figure 53 are pertinent to an alum-primary sludge similar to that produced during Phase 1 thickened to approximately three weight percent and dosed with 37 weight percent lime. Note that a filter submergence of 30 percent was assumed in constructing the full-scale filtration curve and that conditioning chemical was included in the quoted rates.

Use of Figures 53 and 54 for design will be illustrated with the following example. Assume that it would be desired to operate a full-scale vacuum filter to produce a filter cake of 29 weight percent total solids. From Figure 54, a correlating factor

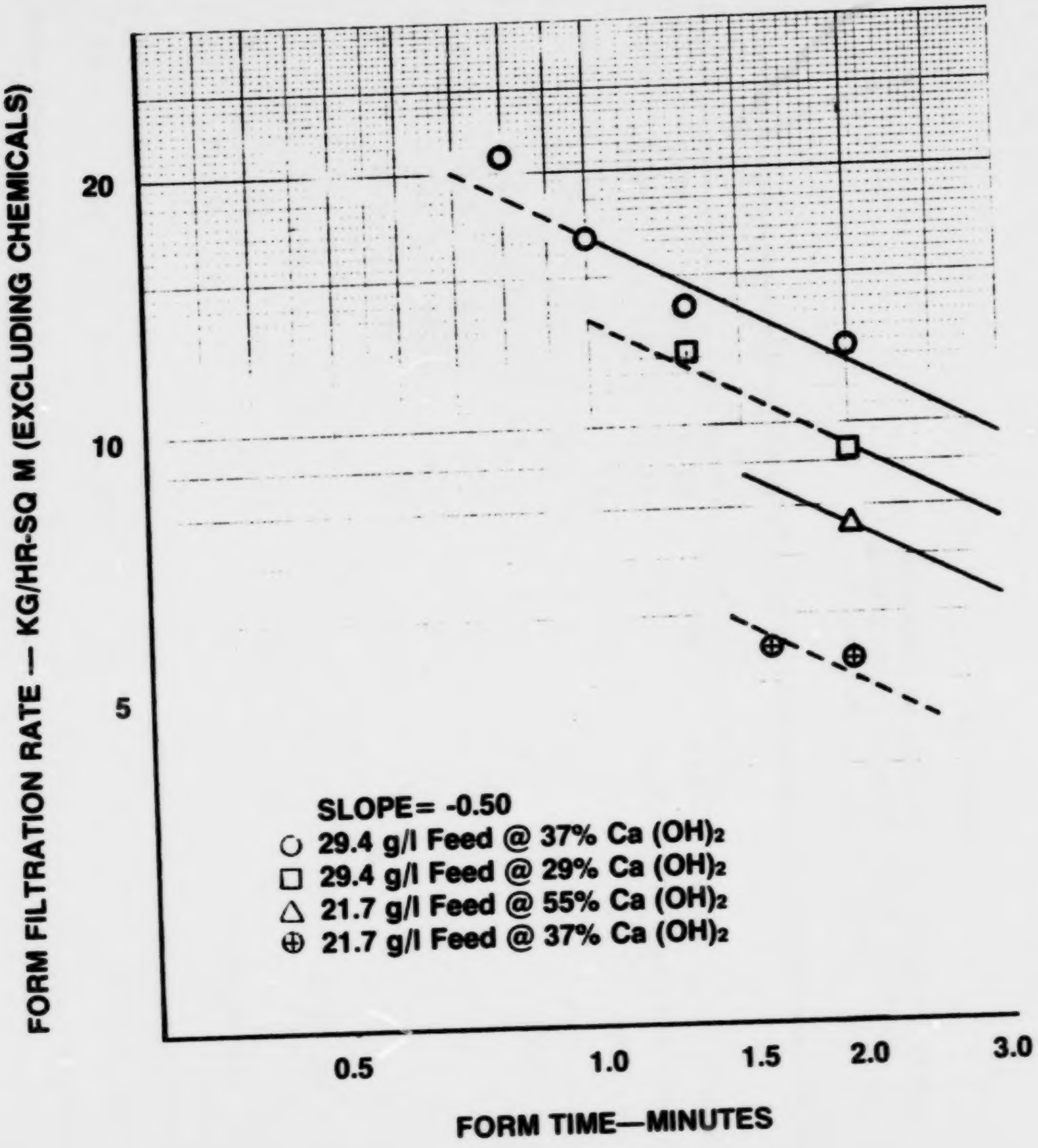

FIGURE 52. FORM FILTRATION RATE VS. FORM TIME FOR ALUM-PRIMARY SLUDGE — PHASE 1.
(High Alum; no Polymer)

TABLE 32

DATA SUMMARY OF VACUUM FILTRATION OF ALUM-PRIMARY SLUDGE FROM PHASE 1

(High Alum; no Polymer)

	RUN NUMBER					
	1	2	3	4	5	6
Feed total solids, g/ℓ	28.4	28.4	21.7	21.7	21.7	29.4
Conditioned feed total solids, g/ℓ	31.0	31.7	24.9	24.5	37.3	27.5
Lime dose, weight percent	24	31	37	37	55	29
Filter form time (θ_f), min.	2.0	2.0	2.0	1.6	2.0	2.0
Filter dry time (θ_d), min.	3.8	3.8	3.8	3.1	3.8	3.8
Form vacuum, mm Hg	640	640	580	580	610	510
Dry vacuum, mm Hg	560	530	530	560	560	460
Discharge code	P	F-P	ND	F	P-F	P
Cake thickness, mm	0.80	0.80	1.2	1.6	1.6	1.2
Cake dry weight (W), Kg/sq m	0.24	0.34	0.24	0.20	0.39	0.40
θ_d/W, Kg-min/sq m	16	11	16	16	10	10
Form filtration rate (Kg/hr-sq m (including chemicals)	7.2	10.2	7.2	7.5	11.7	12.0
Cake solids content, weight percent	28.7	26.2	28.0	25.8	25.9	27.6
Total solids recovery, weight Percent	99	99	98	99	99	97

(continued)

TABLE 32. (continued)

	RUN NUMBER 7	8	9	10	11
Feed total solids, g/ℓ	29.4	29.4	29.4	29.4	29.4
Conditioned feed total solids, g/ℓ	37.7	37.4	38.0	37.8	37.8
Lime dose, weight percent	29	37	37	37	37
Filter form time (θ_f), min.	1.3	1.3	2.0	1.0	0.8
Filter dry time (θ_d), min.	2.6	2.6	3.8	1.9	1.5
Form vacuum, mm Hg	510	510	510	510	510
Dry vacuum, mm Hg	460	460	460	460	510
Discharge code	P	F	F	P-F	P
Cake thickness, mm	1.2	1.2	1.6	1.2	0.80
Cake dry weight (W), Kg/sq m	0.34	0.41	0.56	0.38	0.39
θ_d/W, Kg-min/sq m	7.6	6.3	6.8	5.0	3.9
Form filtration rate (Kg/hr-sq m (including chemicals)	15.7	18.9	16.8	22.8	28.5
Cake solids content, weight percent	27.8	27.8	29.2	28.7	28.3
Total solids recovery, weight Percent	99	97	99	99	97

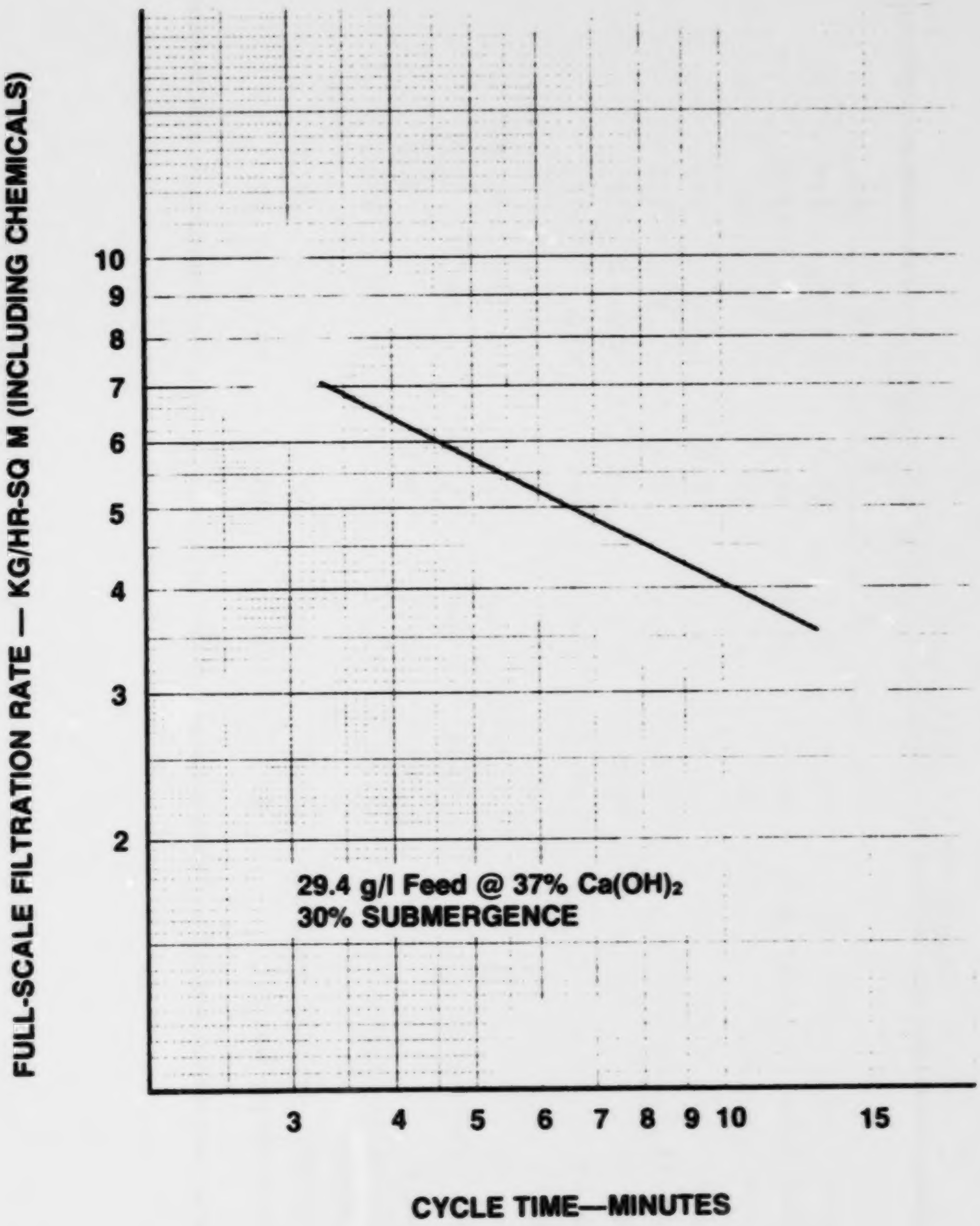

FIGURE 53. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR ALUM-PRIMARY SLUDGE — PHASE 1.
(High Alum; no Polymer)

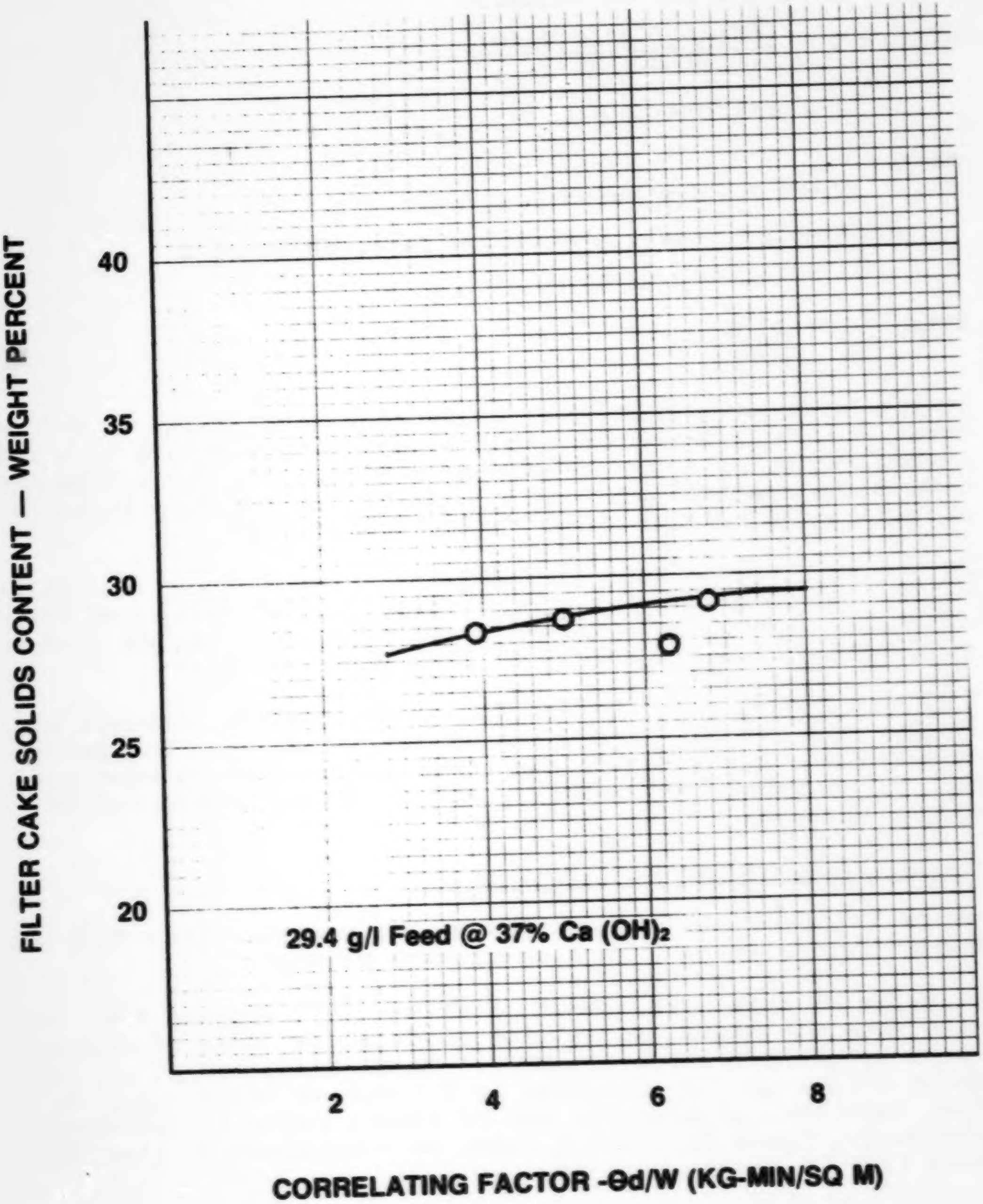

FIGURE 54. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR ALUM-PRIMARY SLUDGE — PHASE 1. (High Alum; no Polymer)

(θ_d/W) of 6.0 would be required

Since, $$W = \frac{(\text{FSFR})\ (\text{Cycle Time})}{60\ \text{min/hr}} \qquad (3)$$

$$\theta_d = 0.5\ (\text{Cycle Time})^* \qquad (4)$$

where:

W = cake dry weight, Kg/sq m

FSFR = full-scale filtration rate, Kg/hr-sq m

θ_d = filter dry time, min

Then, $$\frac{\theta_d}{W} = \frac{30.0}{\text{FSFR}} \qquad (5)$$

In the example, a full-scale filtration rate (FSFR) of 5.0 Kg/hr-sq m (1.0 lb/hr-sq ft) would be calculated with a θ_d/W correlating factor of 6.0. Referring to Figure 53, the filter would have to be operated at a cycle time of 6.5 minutes, which is within the acceptable range for cake discharge.

It should be noted that the FSFR of 5.0 Kg/hr-sq m (1.0 lb/hr-sq ft) includes chemical conditioning. Subtracting out the conditioning lime gives a filter yield rate based on dry sludge feed solids of 3.7 Kg/hr-sq m (0.73 lb/hr-sq ft).

With the exception of volatile suspended solids, capture of feed sludge constituents across the vacuum filter were not determined during Phase 1. Volatile suspended solids capture were evaluated and ranged from 97-99 weight percent with an average of 98 weight percent.

Phase 2 (High Alum; with Polymer)

Table 33 summarizes the results of the vacuum filter runs performed during Phase 2 on alum-primary sludge.

A review of this data, shown on Figure 55, revealed that filtration rate was maximized at lime conditioning dose of approximately 35 weight percent as $Ca(OH)_2$. Tests conducted at a lime dose of 38 weight percent resulted in filtration rates equal to or less than that observed at the 35 weight percent lime dose. Similarly, lower filtration rates were observed at lime doses of less than 35 weight percent.

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*Most belt vacuum filters are constructed such that approximately 50 percent of the cycle time functions as cake dry time. Other designs may be considered by modifying the constant, 0.5, in equation (4).

TABLE 33

DATA SUMMARY OF VACUUM FILTRATION OF
ALUM-PRIMARY SLUDGE FROM PHASE 2
(High Alum; with Polymer)

	RUN NUMBER						
	1	2	3	4	5	6	7
Feed total solids, g/ℓ	28.0	28.0	28.0	28.0	28.0	28.0	28.0
Conditioned feed total solids, g/ℓ	32.7	32.7	32.7	32.7	34.4	34.4	34.4
Lime dose, weight percent	15	15	15	15	30	30	30
Filter form time (θ_f), min.	2.0	1.3	1.0	0.75	2.0	1.3	0.75
Filter dry time, (θ_d), min.	3.8	2.6	1.9	1.5	3.8	2.6	1.5
Form vacuum, mm Hg	510	510	530	530	510	510	530
Dry vacuum, mm Hg	480	480	510	510	480	480	510
Discharge code	E	G-E	G	F	E	G-E	G-E
Cake thickness, mm	1.2	1.2	0.80	0.80	2.5	1.6	1.2
Cake dry weight (W), Kg/sq m	0.37	0.41	0.32	0.24	0.56	0.48	0.38
θ_d/W, Kg-min/sq m	10	6.3	5.9	6.3	6.8	5.4	3.9
Form filtration rate, (Kg.hr-sq m (including chemicals)	11.1	18.9	19.2	19.2	16.8	22.2	30.4
Cake solids content, weight percent	28.0	27.6	27.0	26.0	27.8	27.7	28.0
Total solids recovery, weight percent	99	98	98	94	99	96	99

(continued)

TABLE 33. (continued)

	RUN NUMBER						
	8	9	10	11	12	13	14
Feed total solids, g/ℓ	28.0	29.0	29.0	29.0	29.0	29.0	29.1
Conditioned feed total solid, g/ℓ	34.4	37.3	37.3	37.3	37.3	37.3	34.9
Lime dose, weight percent	30	35	35	35	35	35	18
Filter form time (θ_f), min.	0.59	2.0	1.0	.59	0.45	1.3	2.0
Filter dry time (θ_d), min.	1.2	3.8	1.9	1.2	0.87	2.6	3.8
Form vacuum, mm Hg	530	510	510	530	530	510	510
Dry vacuum, mm Hg	510	480	480	510	510	480	460
Discharge code	G	E	G	F-G	F-P	G-E	F-G
Cake thickness, mm	1.2	2.4	1.6	1.2	0.8	1.6	1.6
Cake dry weight (W), Kg/sq m	0.32	0.77	0.55	0.40	0.34	0.56	0.46
θ_d/W, Kg-min/sq m	3.8	4.9	3.5	3.0	2.6	4.6	8.3
Form filtration rate (Kg/hr-sq m (including chemicals)	32.5	23.1	33.0	40.7	45.3	25.8	13.9
Cake solids content, weight percent	27.9	28.8	28.0	25.7	25.7	72.0	27.0
Total solids recovery, weight percent	98	95	97	99	98	99	99

(continued)

TABLE 33. (continued)

	RUN NUMBER						
	15	16	17	18	19	20	21
Feed total solids, g/ℓ	29.1	29.1	29.1	29.1	29.1	29.1	29.1
Conditioned feed total solids, g/ℓ	34.9	37.3	37.7	36.6	36.9	37.1	37.1
Lime dose, weight percent	18	25	25	25	25	38	38
Filter form time (θ_f), min.	1.3	1.3	1.0	0.75	2.0	2.0	1.0
Filter dry time (θ_d), min.	2.6	2.6	1.9	1.5	3.8	3.8	1.9
Form vacuum, mm Hg	510	510	510	510	510	510	510
Dry vacuum, mm Hg	480	480	480	480	480	480	480
Discharge code	F-G	F-G	F	F-P	G	G-E	F-G
Cake thickness, mm	0.8	1.6	1.6	0.8	1.6	3.2	1.6
Cake dry weight (W), Kg/sq m	0.36	0.53	0.47	0.37	0.56	0.71	0.49
θ_d/W, Kg-min/sq m	7.2	4.9	4.0	4.1	6.8	5.4	3.9
Form filtration rate (Kg/hr-sq m (including chemicals)	16.6	24.5	28.2	29.6	16.8	21.3	29.4
Cake solids content, weight percent	28.0	27.0	26.4	26.3	27.4	27.7	27.4
Total solids recovery, weight percent	98	96	96	93	98	98	96

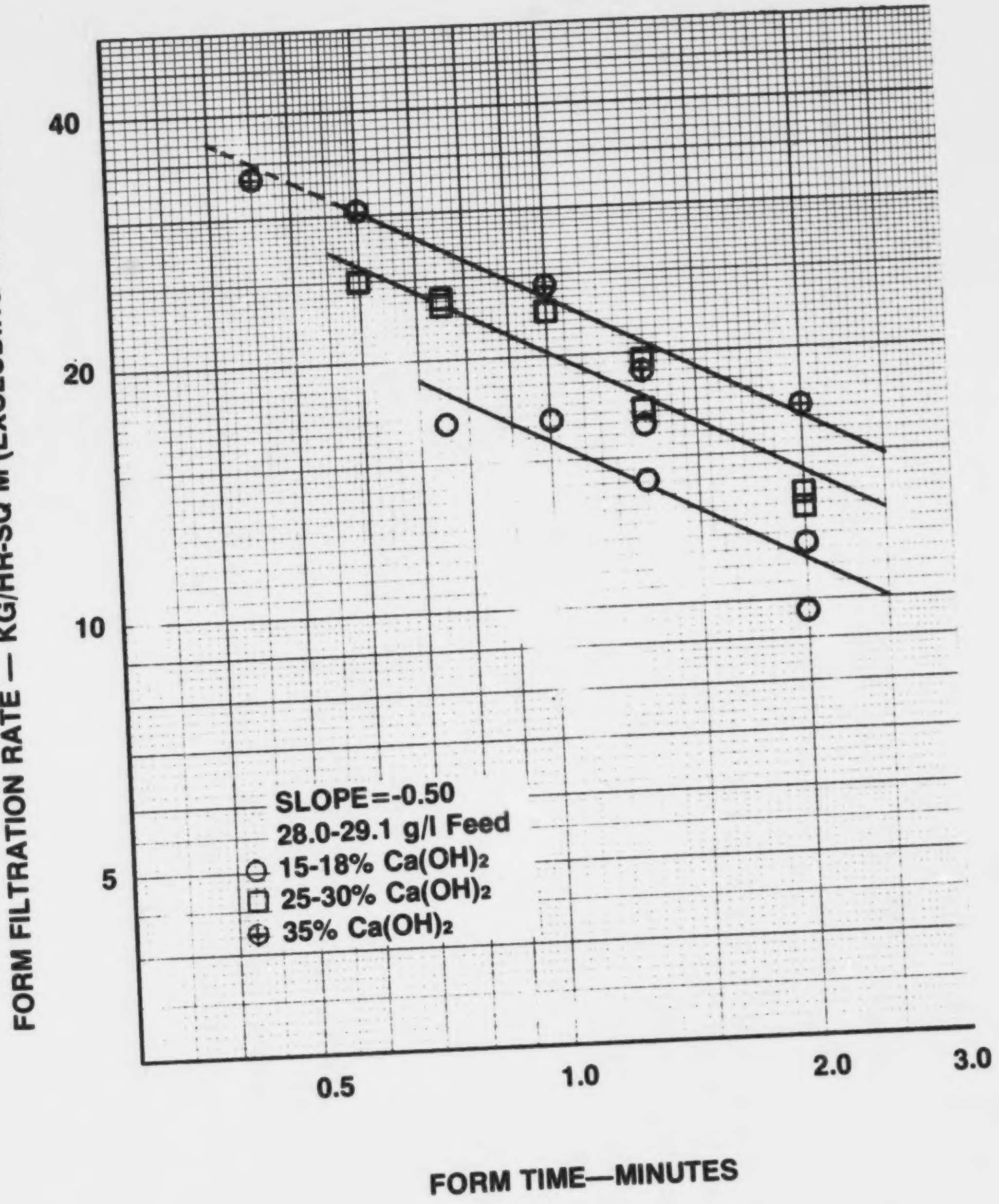

FIGURE 55. FORM FILTRATION RATE VS. FORM TIME FOR ALUM-PRIMARY SLUDGE — PHASE 2.
(High Alum; with Polymer)

Figures 56 and 57 summarize the full-scale operating curves for two levels of lime conditioning. Use of these curves is identical to the description contained in the previous section. The data shown on Figure 57 indicate that cake dry solids concentration was independent of lime conditioning dose. This is contrary to what was expected. It was hypothesized that the cake was at near maximum solids concentration due to the thin cakes 0.8-3.2 mm (0.3-0.13 inch) produced and, therefore, the effect of the lime component was minimal.

Capture of the volatile suspended solids and aluminum constituents of the filter feed sludge were determined. Volatile solids capture ranged from 93-99 weight percent, with an average of 98 weight percent. Aluminum capture averaged 88 weight percent.

Phase 3 (Low Alum; with Polymer)

Table 34 summarizes the vacuum filter data developed from operation on the alum-primary sludge generated during Phase 3.

Figure 58 shows the vacuum filter results for the five operating conditions evaluated. From Figure 58, it can be seen that form filtration rate increased as the lime conditioning dose increased from 15 weight percent to 26 weight percent over a feed solids concentration range of 47.2 to 61.4 g/ℓ TS. The data also demonstrated that increases in feed TS concentration resulted in increases in form filtration rate at constant lime conditioning dose.

From Table 34, the results of runs 12-15 indicate that increasing the lime dose to 35 weight percent resulted in lower filtration rates than those shown on Figure 58 for 25 weight percent lime addition at feed concentrations ranging from 47.2 to 61.4 g/ℓ TS. With lower feed concentrations, 40.6 g/ℓ TS, increasing the lime dose to 37 weight percent may be deduced to have resulted in an increase in filtration rate.

Figures 59 and 60 show the predicted full-scale filtration performance for the four operating conditions judged to be most typical for the alum-primary sludge generated in Phase 3. As observed in Phase 2, filter cake solids content was found to be independent of feed sludge concentration and lime conditioning dose.

Phosphorus, volatile suspended solids and aluminum constituent captures were determined during Phase 3. Volatile suspended solids capture ranged from 96 to 99 weight percent with an average of 98 weight percent. Both phosphorus and aluminum capture averaged 97 weight percent.

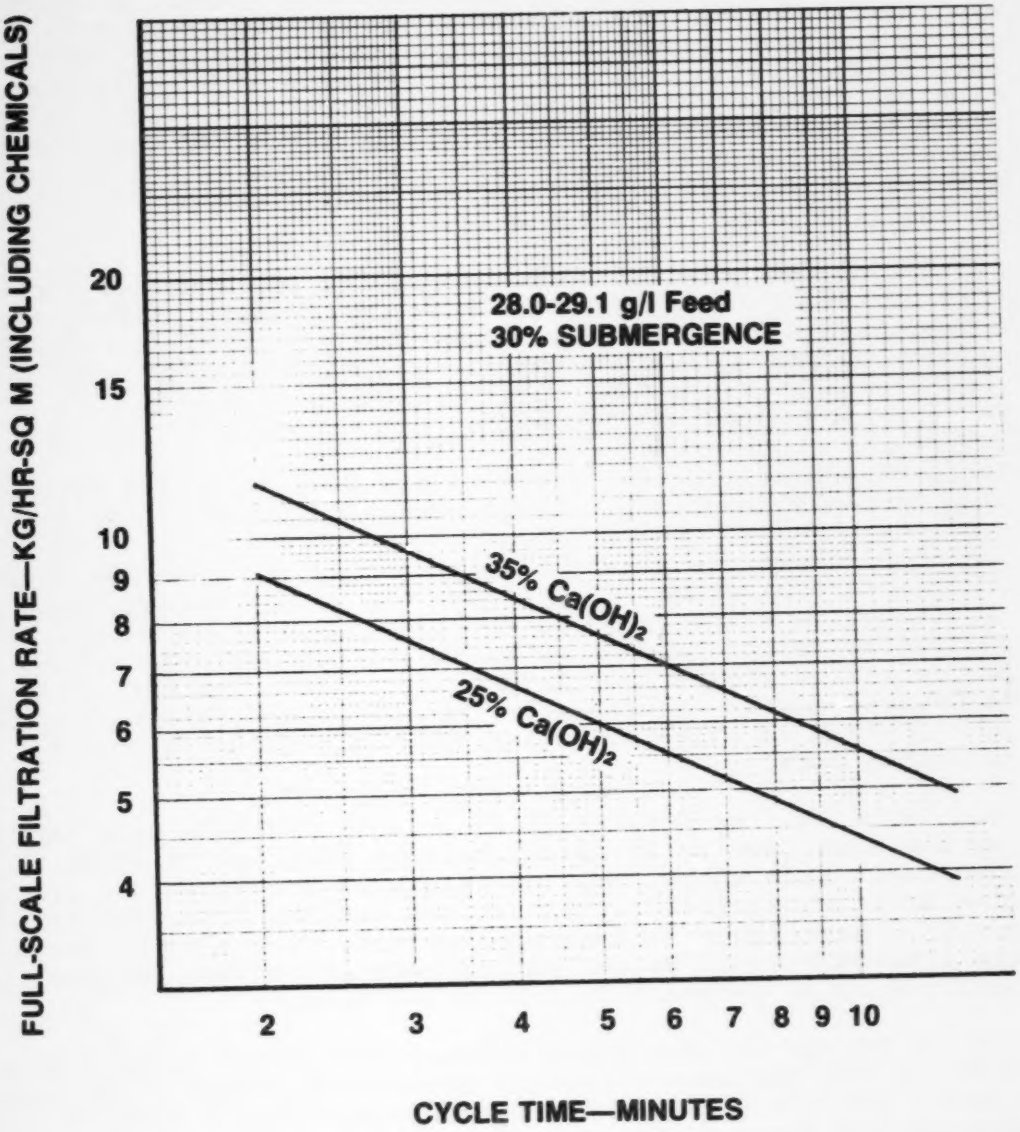

FIGURE 56. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR ALUM-PRIMARY SLUDGE — PHASE 2.
(High Alum; with Polymer)

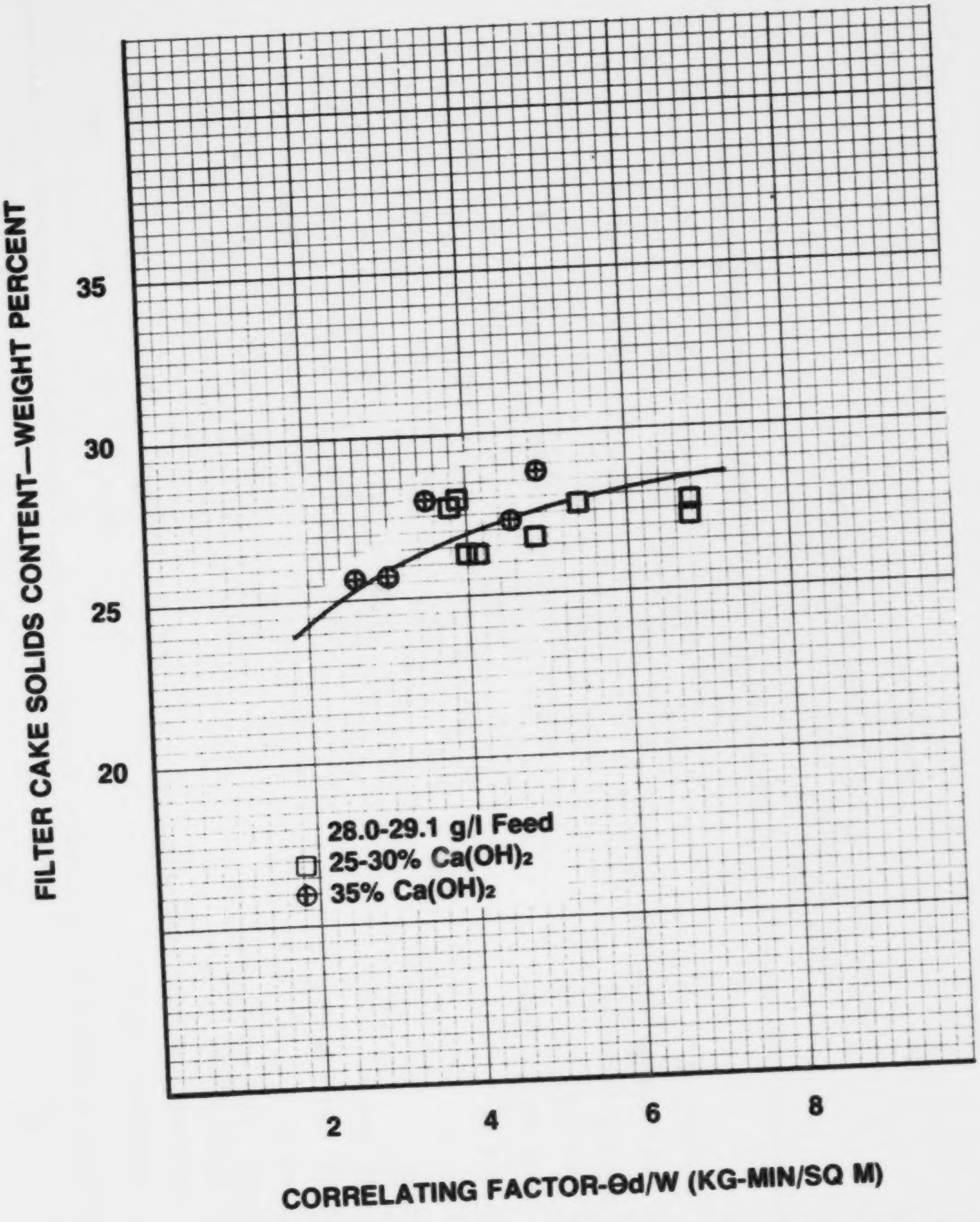

FIGURE 57. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR ALUM-PRIMARY SLUDGE — PHASE 2. (High Alum; with Polymer)

TABLE 34

DATA SUMMARY OF VACUUM FILTRATION OF ALUM-PRIMARY SLUDGE FROM PHASE 3

(Low Alum; with Polymer)

	RUN NUMBER					
	1	2	3	4	5	6
Feed total solids, g/ℓ	40.6	40.6	40.6	40.6	47.2	47.2
Conditioned feed total solids, g/ℓ	53.2	53.2	53.2	53.2	48.8	48.8
Lime dose, weight percent	37	37	37	37	15	15
Filter form time (θ_f), min.	2.0	1.0	0.59	0.41	2.0	1.0
Filter dry time (θ_d), min.	3.8	1.9	1.2	0.80	3.8	1.9
Form vacuum, mm Hg	510	510	510	510	510	510
Dry vacuum, mm Hg	460	460	480	480	460	430
Discharge code	E	E	G	F	G	F-G
Cake thickness, mm	4.8	3.2	2.4	1.6	2.4	1.6
Cake dry weihgt (W), Kg/sq m	1.3	0.96	0.73	0.59	0.64	0.47
θ_d/W, Kg-min/sq m	2.9	2.0	1.6	1.4	5.9	4.0
Form filtration rate (Kg/hr-sq m (including chemicals)	39.0	57.6	74.2	86.3	19.2	28.2
Cake solids content, weight percent	26.3	25.8	26.2	26.2	24.8	25.2
Total solids recovery, weight percent	99	99	99	98	99	99

(continued)

TABLE 34. (continued)

	RUN NUMBER					
	7	8	9	10	11	12
Feed total solids, g/ℓ	47.2	47.2	47.2	47.2	47.2	57.7
Conditioned feed total solids, g/ℓ	48.8	51.6	51.6	51.6	51.6	68.8
Lime dose, weight percent	15	25	25	25	25	35
Filter form time (θ_f), min.	0.75	1.0	0.75	0.59	2.0	2.0
Filter dry time (θ_d), min.	1.5	1.9	1.5	1.2	2.0	2.0
Form vacuum, mm Hg	510	530	530	530	530	510
Dry vacuum, mm Hg	460	480	480	460	480	450
Discharge code	F-P	G	G	F	E	G
Cake thickness, mm	1.2	2.4	2.0	1.6	3.2	1.6
Cake dry weight (W), Kg/sq m	0.38	0.63	0.54	0.52	0.89	0.79
θ_d/W, Kg-min/sq m	3.9	3.0	2.8	2.3	4.3	4.8
Form filtration rate (Kg/hr-sq m (including chemicals)	30.4	37.8	43.2	52.9	26.7	23.7
Cake solids content, weight percent	25.2	26.0	25.8	24.0	25.6	35.4
Total solids recovery, weight percent	98	98	96	98	97	99

(continued)

TABLE 34. (continued)

	RUN NUMBER					
	13	14	15	16	17	18
Feed total solids, g/ℓ	57.7	57.7	57.7	61.4	61.4	61.4
Conditioned feed total solids, g/ℓ	68.8	68.8	68.8	69.1	69.1	69.1
Lime dose, weight percent	35	35	35	16	16	16
Filter form time (θ_f), min.	1.0	0.75	1.3	2.0	1.0	0.59
Filter dry time (θ_d), min.	1.9	1.5	2.6	3.8	1.9	1.2
Form vacuum, mm Hg	510	510	510	510	510	510
Dry vacuum, mm Hg	450	460	430	460	460	460
Discharge code	F	ND	F-G	E	F-G	P-F
Cake thickness, mm	1.2	0.8	1.2	3.2	2.4	1.6
Cake dry weight (W), Kg/sq m	0.50	0.43	0.56	1.1	0.76	0.59
θ_d/W, Kg-min/sq m	3.8	3.5	4.6	3.5	2.5	2.0
Form filtration rate (Kg/hr-sq m (including chemicals)	30.0	34.4	25.8	33.0	45.6	60.0
Cake solids content, weight percent	35.6	34.1	35.1	26.3	27.6	26.7
Total solids recovery, weight percent	99	-	99	99	99	99

(continued)

TABLE 34. (continued)

	RUN NUMBER					
	19	20	21	22	23	24
Feed total solids, g/ℓ	61.4	61.4	61.4	61.4	61.4	61.4
Conditioned feed total solids, g/ℓ	69.1	70.0	70.0	70.0	70.0	70.0
Lime dose, weight percent	16	26	26	26	26	26
Filter form time (θ_f), min.	1.3	2.0	1.0	0.59	0.41	0.32
Filter dry time (θ_d), min.	2.6	3.8	1.9	1.2	0.80	0.62
Form vacuum, mm Hg	510	510	510	510	510	510
Dry vacuum, mm Hg	460	480	460	460	460	480
Discharge code	G	E	E	G	F	P
Cake thickness, mm	2.8	4.8	4.0	3.2	2.4	1.6
Cake dry weight (W), Kg/sq m	0.80	1.4	1.1	0.80	0.67	0.59
θ_d/W, Kg-min/sq m	3.3	2.7	1.7	1.5	1.2	1.1
Form filtration rate (Kg/hr-sq m (including chemicals)	36.9	42.0	66.0	81.4	98.0	111
Cake solids content, weight percent	26.8	26.0	24.6	25.0	25.0	25.0
Total solids recovery, weight percent	98	99	99	99	98	98

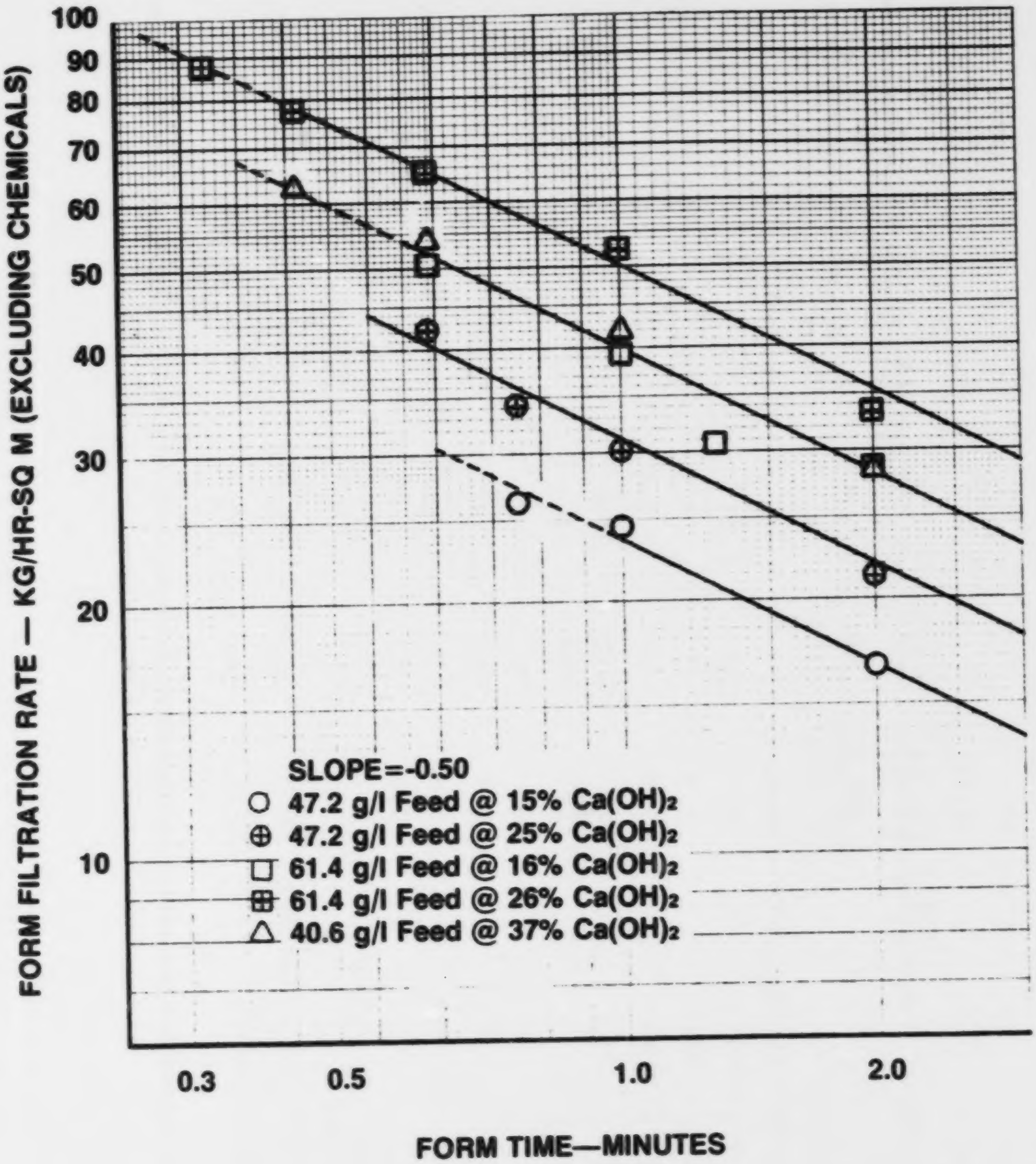

FIGURE 58. FORM FILTRATION RATE VS. FORM TIME FOR ALUM-PRIMARY SLUDGE — PHASE 3.
(Low Alum; with Polymer)

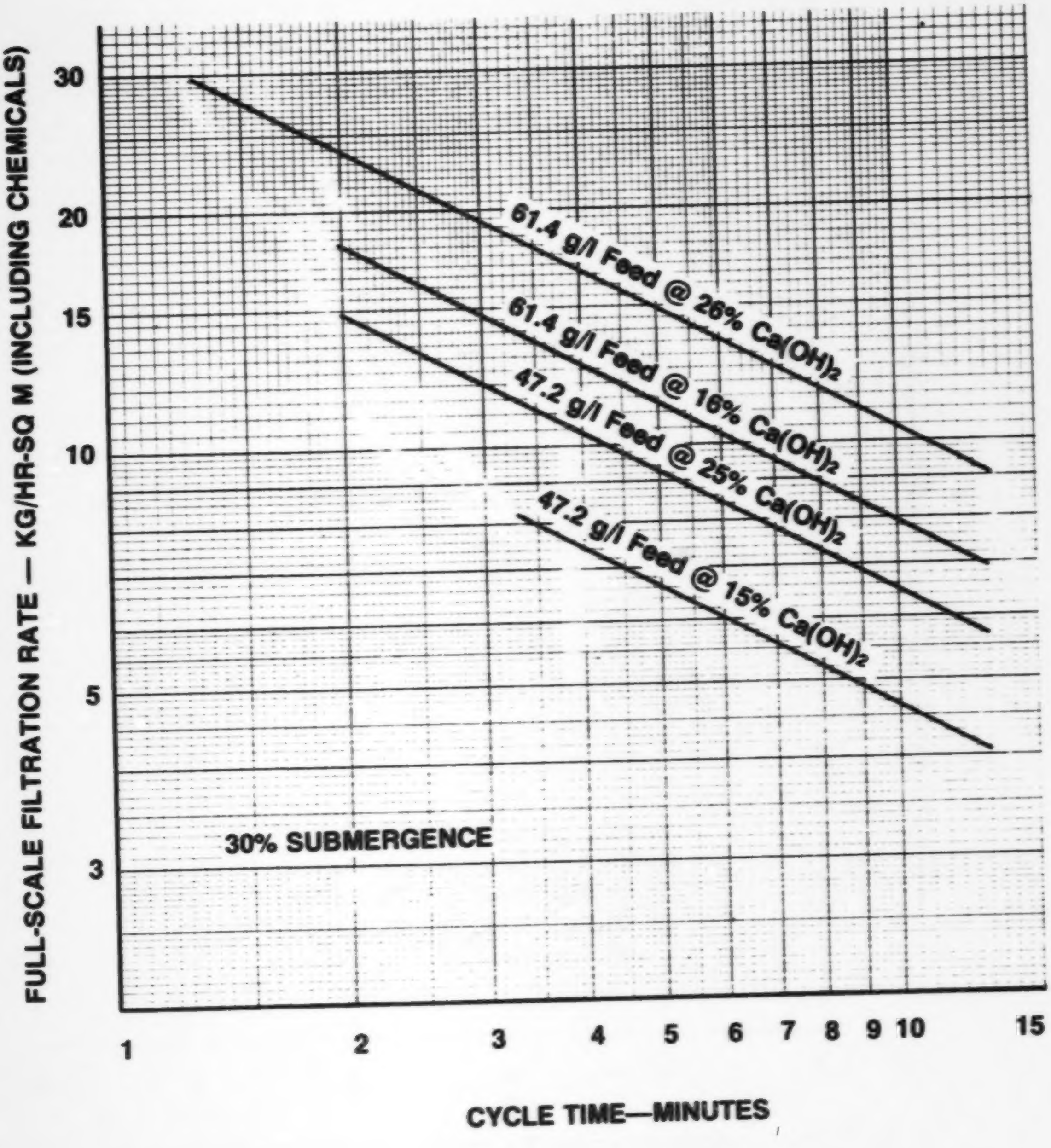

FIGURE 59. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR ALUM-PRIMARY SLUDGE — PHASE 3.
(Low Alum; with Polymer)

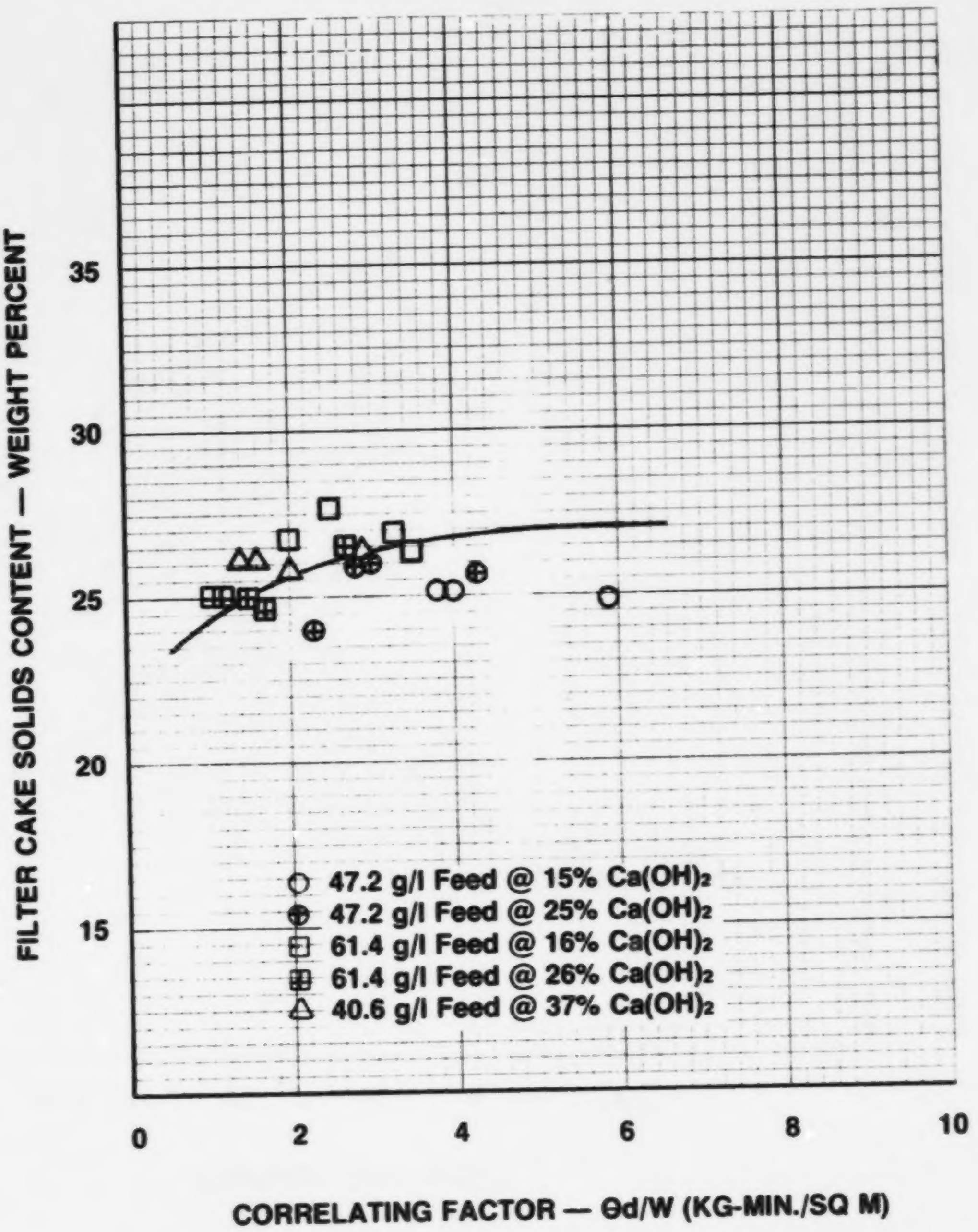

FIGURE 60. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR ALUM-PRIMARY SLUDGE — PHASE 3. (Low Alum; with Polymer)

Phase 4 (High Alum; with Polymer)

A summary of the filter runs conducted on the alum-primary sludge generated during Phase 4 is shown in Tab[illegible]5.

A review of this data, shown on Figures [illegible]ndicated that form filtration rates were maximized at a lime [illegible]itioning dose of approximately 26 weight percent as Ca(OH)[illegible]

The full-scale vacuum filter performance [illegible]s are plotted on Figures 62 and 63. As was observed in al[illegible]vious runs, filter cake solids content was unaffected by lime [illegible]ditioning dose within the range studied.

Phosphorus, volatile suspended solids and [illegible]minum captures were determined during Phase 4. Volatile susp[illegible] solids capture ranged from 94 to 99 weight percent. Pho[illegible]us and aluminum captures determined from results of one va[illegible] filter test series indicated average captures of 80 an[illegible]weight percent, respectively.

FERRIC-PRIMARY SLUDGE

Phase 1 (High Ferric; no Polymer)

Table 36 represents the operating data developed for vacuum filtration of ferric-primary sludge generated during Phase 1.

Figure 64 shows a graphical summary of the three operating conditions studied during this Phase. As indicated, the -0.55 slope of the logarithmic plot of form filtration rate versus form time was only slightly in excess of the theoretical -0.5 slope.

The data on Figure 64 indicate an increase in form filtration rate due to an increase in lime conditioning dose from 18 percent to 30 percent. A further rate increase was observed with a lime conditioning dose increase to 47 percent, however, the full extent of the rate increase was confused due to vacuum fluctuations which were experienced at this lime dose.

Severe filter cake cracking during the cake drying cycle was the cause of the vacuum fluctuations experienced at the 47 percent lime conditioning dose. The diminution in vacuum level also resulted in wetter filter cakes. A review of the data in Table 36 indicated that the vacuum diminution resulted in two to eight percentage points decrease in filter cake dry solids content as compared to filter operation at 30 percent lime. The conclusion was, therefore, that filter operation at a lime conditioning dose of 47 percent would not be advisable.

TABLE 35

DATA SUMMARY OF VACUUM FILTRATION OF ALUM-PRIMARY SLUDGE FROM PHASE 4
(High Alum; with Polymer)

	RUN NUMBER				
	1	2	3	4	5
Feed total solids, g/ℓ	55.2	55.2	55.2	55.2	57.4
Conditioned feed total solids, g/ℓ	66.4	66.4	66.4	66.4	64.9
Lime dose, weight percent	35	35	35	35	26
Filter form time (θ_f), min.	2.0	0.75	0.41	1.0	2.0
Filter dry time (θ_d), min.	3.8	1.5	0.80	1.9	3.8
Form vacuum, mm Hg	510	510	510	510	530
Dry vacuum, mm Hg	460	460	460	460	510
Discharge code	G-E	F-G	P-F	G	G-E
Cake thickness, mm	3.2	2.0	1.2	2.4	3.2
Cake dry weight (W), Kg/sq m	0.93	0.60	0.42	0.67	0.80
θ_d/W, Kg-min/sq m	4.1	2.5	1.9	2.8	4.8
Form filtration rate (Kg/hr-sq m (including chemicals)	27.9	48.0	61.5	40.2	24.0
Cake solids content, weight percent	28.4	27.0	26.7	26.7	27.2
Total solids recovery, weight percent	98	97	89	92	99

(continued)

TABLE 35. (continued)

	RUN NUMBER		
	6	7	8
Feed total solids, g/ℓ	57.4	57.4	57.4
Conditioned feed total solids, g/ℓ	64.9	64.9	64.9
Lime dose, weight percent	26	26	26
Filter form time (θ_f), min.	1.0	0.59	0.75
Filter dry time (θ_d), min.	1.9	1.1	1.5
Form vacuum, mm Hg	510	510	530
Dry vacuum, mm Hg	480	460	480
Discharge code	F-G	F-P	F-G
Cake thickness, mm	1.6	1.2	1.6
Cake dry weight (W), Kg/sq m	0.58	0.45	0.47
θ_d/W, Kg-min/sq m	3.3	2.4	3.2
Form filtration rate (Kg/hr-sq m (including chemicals)	34.8	45.8	37.6
Cake solids content, weight percent	26.8	25.8	25.3
Total solids recovery, weight percent	99	90	90

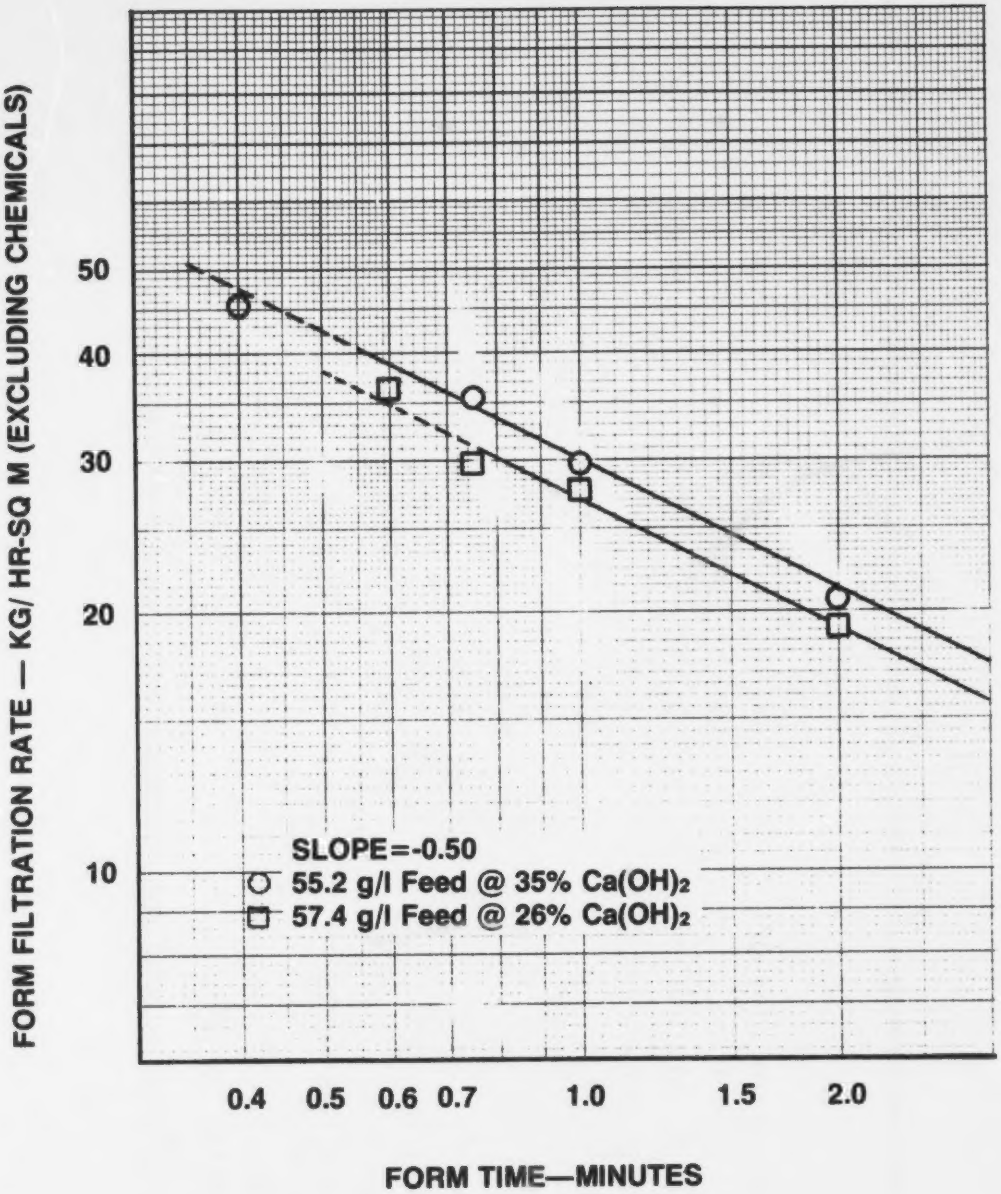

FIGURE 61. FORM FILTRATION RATE VS. FORM TIME FOR ALUM-PRIMARY SLUDGE — PHASE 4.
(High Alum; with Polymer)

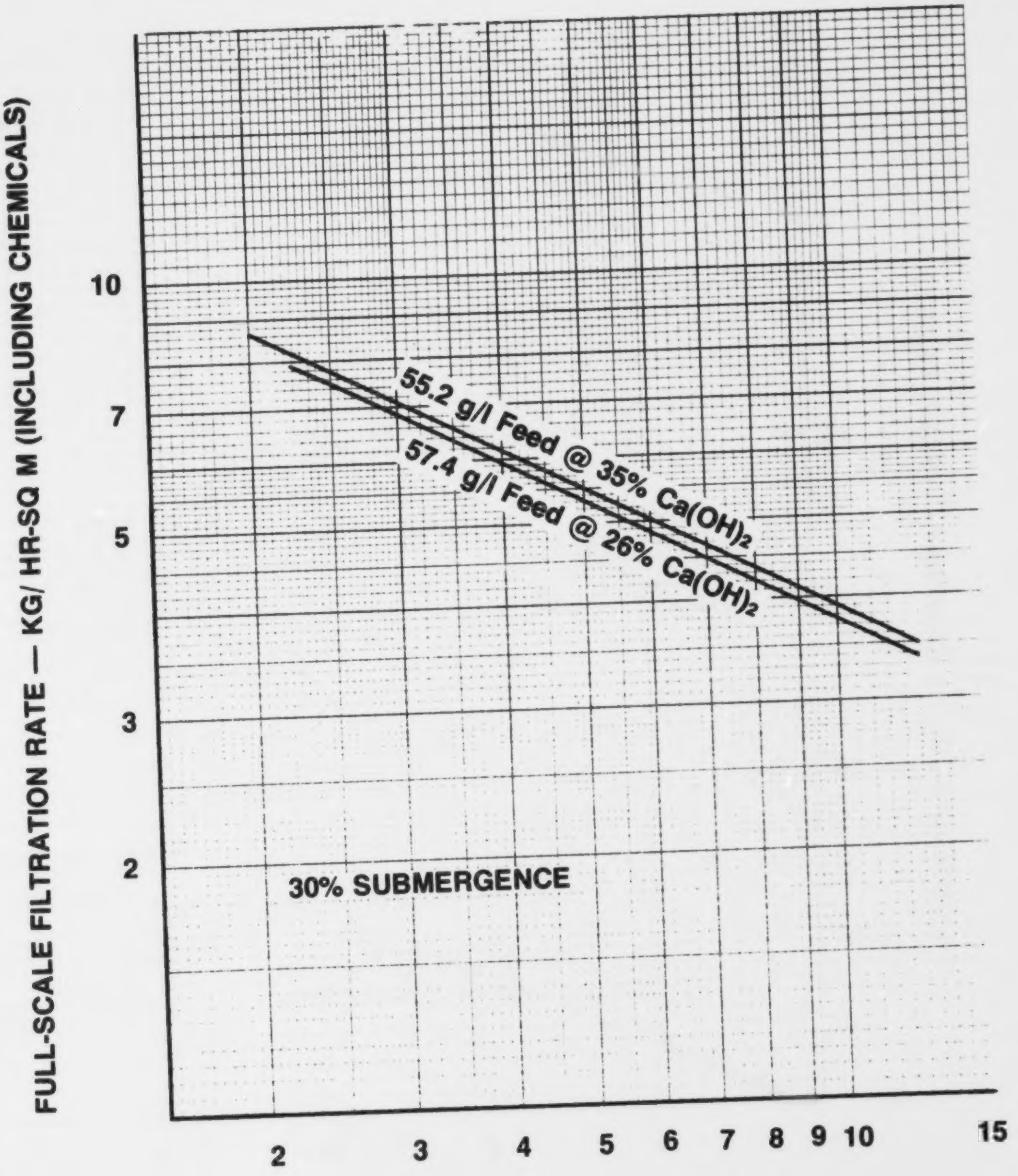

FIGURE 62. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR ALUM-PRIMARY SLUDGE — PHASE 4.
(High Alum; with Polymer)

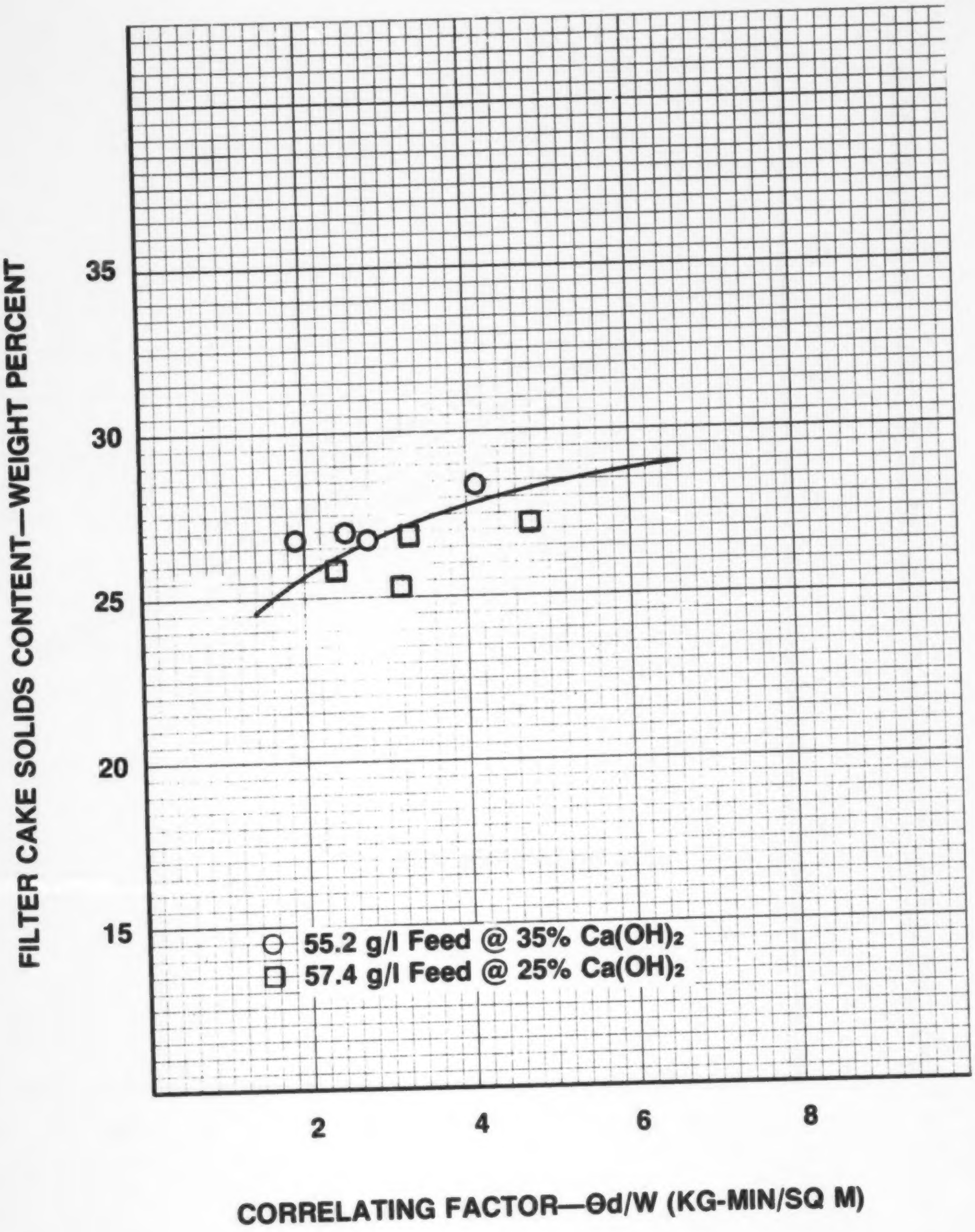

FIGURE 63. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR ALUM-PRIMARY SLUDGE — PHASE 4. (High Alum; with Polymer)

TABLE 36

DATA SUMMARY OF VACUUM FILTRATION OF FERRIC-PRIMARY SLUDGE FROM PHASE 1
(High Ferric; no Polymer)

	RUN NUMBER						
	1	2	3	4	5	6	7
Feed total solids, g/ℓ	41.2	41.2	41.2	41.2	46.5	46.5	46.5
Conditioned feed total solids, g/ℓ	48.1	49.6	50.0	50.5	47.2	47.2	47.2
Lime dose, weight percent	47	47	47	47	18	18	18
Filter form time (θ_f), min.	2.0	1.0	0.59	0.41	2.0	1.3	1.6
Filter dry time (θ_d), min.	3.8	1.9	1.2	0.80	3.8	2.6	3.1
Form vacuum, mm Hg	250	330	410	470	580	580	580
Dry vacuum, mm Hg	230	250	360	420	510	510	510
Discharge code	E	E	G	F-G	G	F	F
Cake thickness, mm	5.6	3.6	2.4	1.6	1.6	1.2	1.2
Cake dry weight (W), Kg/sq m	1.4	1.0	0.83	0.68	0.58	0.44	0.49
θ_d/W, Kg-min/sq m	2.7	1.9	1.5	1.2	6.6	5.9	6.3
Form filtration rate (Kg/hr-sq m (including chemicals)	42.0	60.0	84.4	99.5	17.4	20.3	18.4
Cake solids content, weight percent	26.1	28.2	29.6	31.7	27.8	29.5	30.5
Total solids recovery, weight percent	99	99	99	98	97	98	98

(continued)

TABLE 36. (continued)

	RUN NUMBER						
	8	9	10	11	12	13	14
Feed total solids, g/ℓ	46.5	46.5	46.5	46.5	46.5	46.5	46.5
Conditioned feed total solids, g/ℓ	47.2	48.2	48.2	48.2	48.2	48.2	48.2
Lime dose, weight percent	18	30	30	30	30	30	30
Filter form time (θ_f), min.	1.1	2.0	1.6	1.1	2.0	1.0	1.6
Filter dry time (θ_d), min.	2.2	3.8	3.1	2.2	3.8	1.9	3.1
Form vacuum, mm Hg	580	580	580	580	580	580	580
Dry vacuum, mm Hg	510	510	510	510	510	510	510
Discharge code	P	G	G	F	G-E	G	G
Cake thickness, mm	0.80	2.4	2.0	2.0	2.4	1.6	2.4
Cake dry weight (W), Kg/sq m	0.39	.68	0.67	0.63	0.78	0.58	0.59
θ_d/W, Kg-min/sq m	5.6	5.6	4.6	3.5	4.9	3.3	5.3
Form filtration rate (Kg/hr-sq m (including chemicals)	21.3	20.4	25.1	34.4	23.4	34.8	22.1
Cake solids content, weight percent	31.7	33.7	33.6	33.7	31.4	34.1	34.0
Total solids recovery, weight percent	99	-	99	99	99	99	99

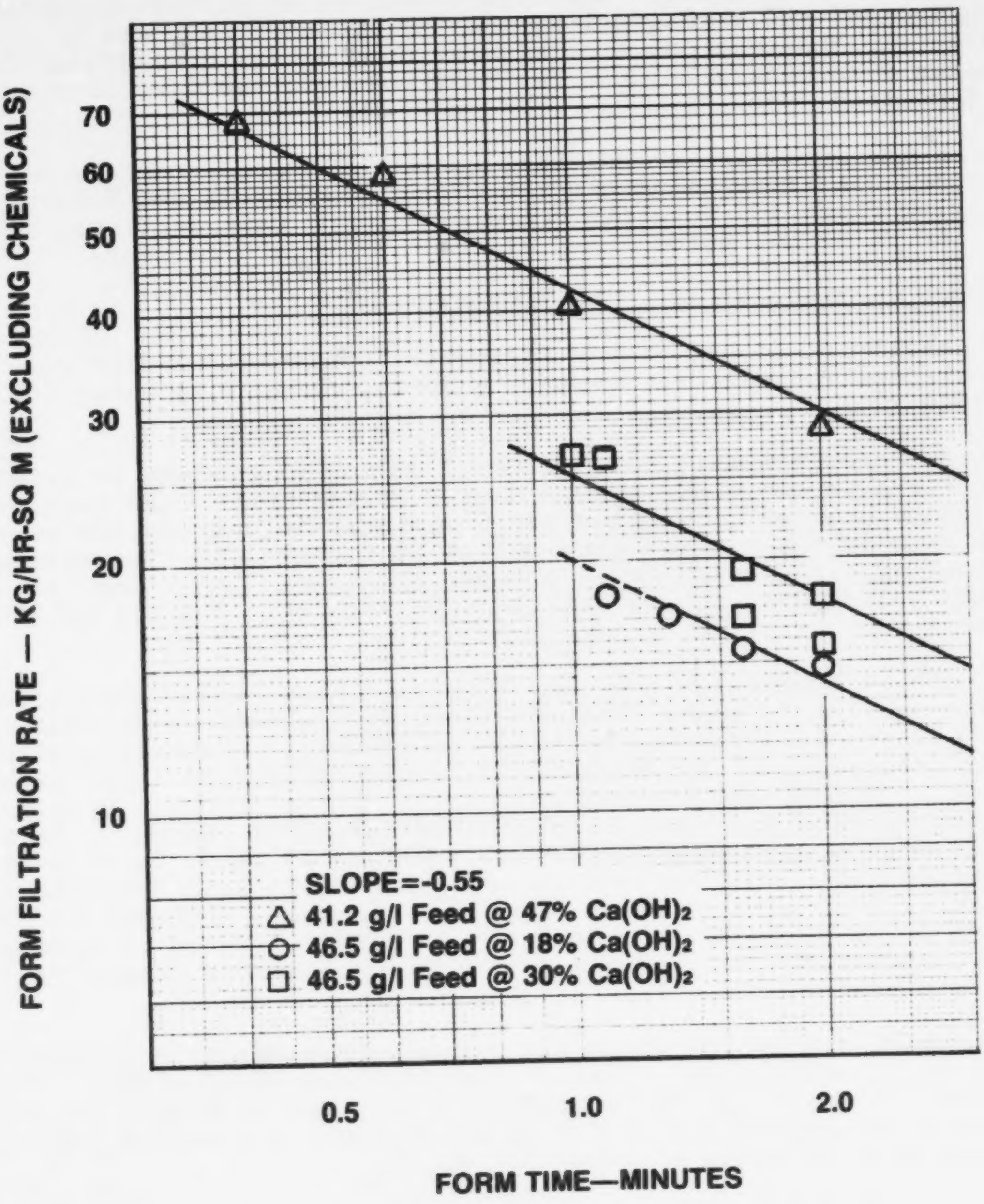

FIGURE 64. FORM FILTRATION RATE VS FORM TIME FOR FERRIC-PRIMARY SLUDGE — PHASE 1. (High Ferric; no Polymer)

Figures 65 and 66 show the full-scale vacuum filter operating curves for the ferric-primary sludge produced in Phase 1. Unlike the observations from alum-primary sludge dewatering, filter cakes were generally drier at a 30 percent lime dose than at an 18 percent dose.

With the exception of volatile suspended solids, capture of feed sludge constituents across the vacuum filter was not determined during Phase 1. Volatile suspended solids capture ranged from 97 to 99 weight percent and averaged 98 weight percent.

Phase 2 (High Ferric; with Polymer)

A tabulation of the vacuum filtration operating results for ferric-primary sludge dewatering during Phase 2 is shown in Table 37.

Filter cake cracking also occurred during this phase at lime conditioning doses above approximately 20 percent. This resulted in vacuum level diminution and a decrease in cake solids content of zero to three percentage points as compared to the cakes produced at either a 10 to 20 percent lime dose. No effect on cake solids content was observed at lime doses from 10 to 20 percent.

Figure 67 summarizes the cake formation rate versus form time data developed during Phase 2. At a feed sludge concentration of 43.6 to 44.6 g/ℓ TS, benefits in form filtration rate were not observed with an increase in lime dose from 10 to 20 percent. An increase in form filtration rate did occur, however, when the lime dose was increased further to 24 percent. At a feed sludge concentration of 53.3 g/ℓ, an increase in form filtration rate was observed with an increase in lime conditioning dose from 10 percent to 19 percent. A further increase in lime dose to 26 percent (at a 60.2 g/ℓ feed TS) resulted in erratic performance due to vacuum fluctuations which were caused by cake cracking.

Figure 68 summarizes operating results from the more important operating conditions of Phase 2 on ferric-primary sludge. These curves, in addition to the information shown on Figure 69, represent full-scale vacuum filter performance and may be used for predictive purposes as explained above.

Capture of the volatile suspended solids, phosphorus and iron constituents of the filter feed sludge were determined. Volatile suspended solids capture ranged from 95 to 99 weight percent with an average of 98 weight percent. Both phosphorus and iron capture averaged 99 weight percent.

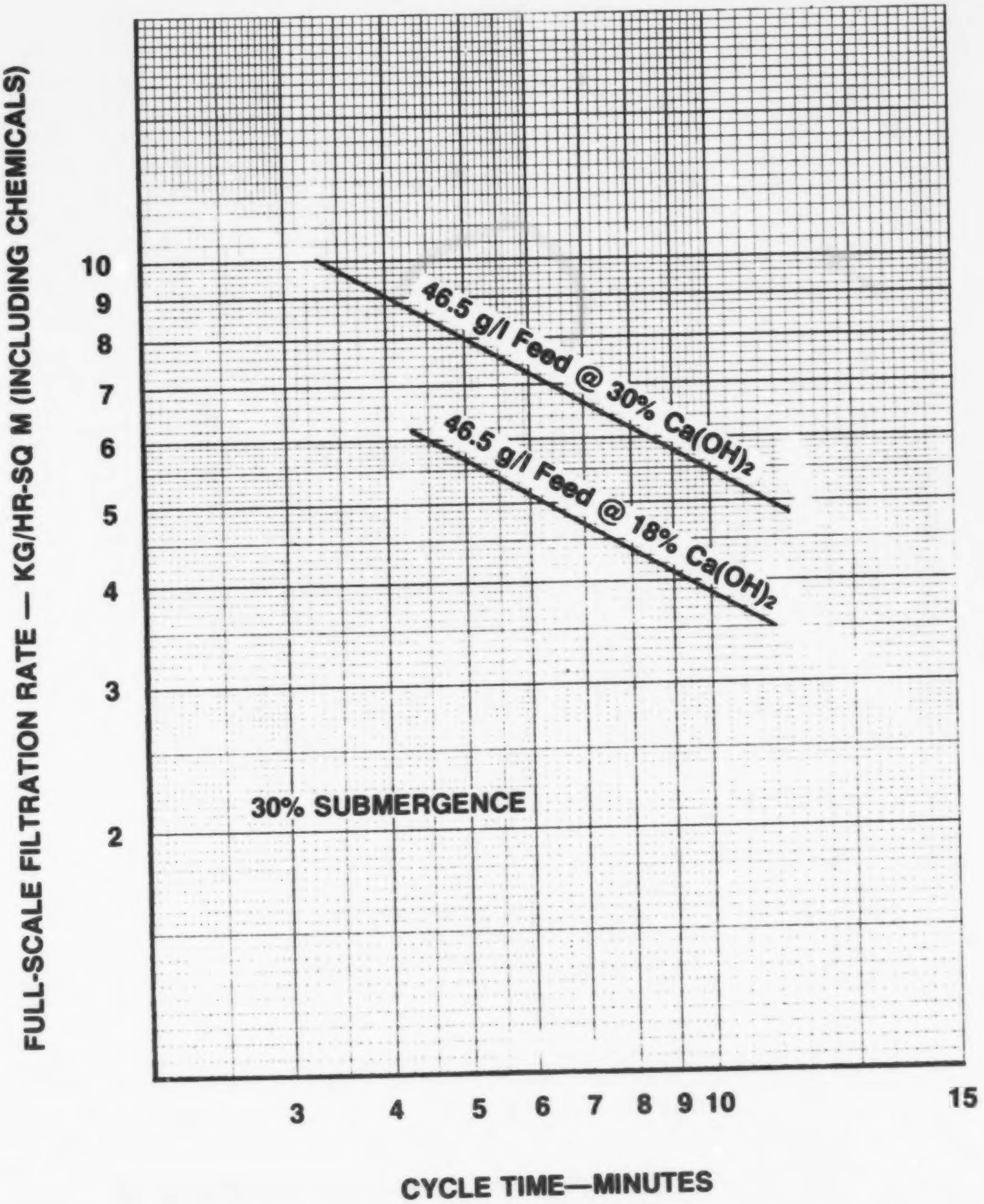

FIGURE 65. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR FERRIC-PRIMARY SLUDGE — PHASE 1.
(High Ferric; no Polymer)

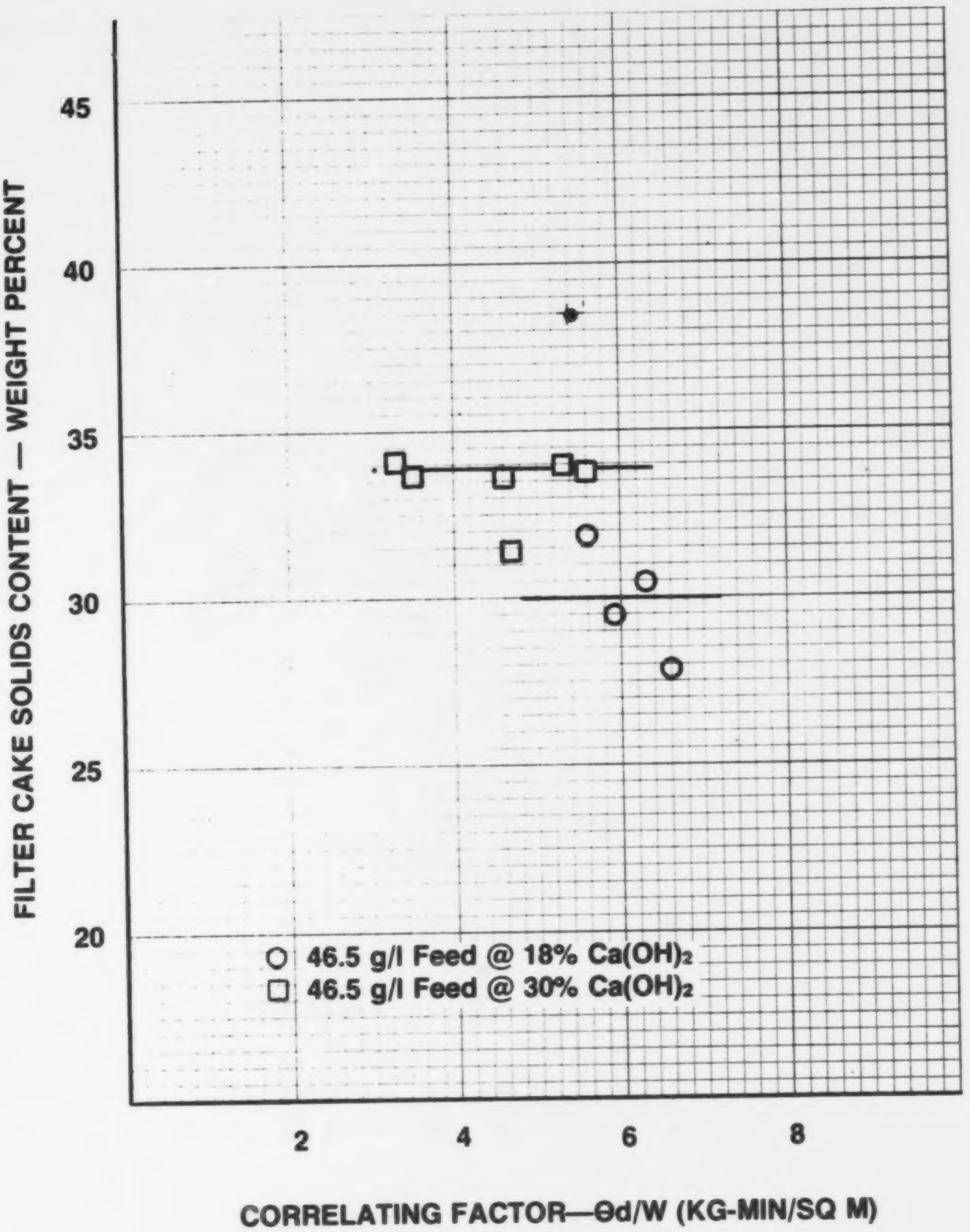

FIGURE 66. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR FERRIC-PRIMARY SLUDGE — PHASE 1.
(High Ferric; no Polymer)

TABLE 37

DATA SUMMARY OF VACUUM FILTRATION OF FERRIC-PRIMARY SLUDGE FROM PHASE 2
(High Ferric; with Polymer)

	RUN NUMBER					
	1	2	3	4	5	6
Feed total solids, g/ℓ	43.6	43.6	43.6	43.6	44.6	44.6
Conditioned feed total solids, g/ℓ	49.3	49.3	49.3	49.3	45.5	45.5
Lime dose, weight percent	24	24	24	24	10	10
Filter form time (θ_f), min.	2.0	0.41	0.59	1.0	2.0	1.0
Filter dry time (θ_d), min.	3.8	0.80	1.2	1.9	3.8	1.9
Form vacuum, mm Hg	250	460	410	330	510	530
Dry vacuum, mm Hg	200	360	330	280	480	510
Discharge code	E	F-P	F-G	G-E	E	G-E
Cake thickness, mm	4.8	1.6	2.0	3.2	3.2	1.6
Cake dry weight (W), Kg/sq m	1.4	0.63	0.75	0.95	0.87	0.60
θ_d/W, Kg-min/sq m	2.7	1.3	1.6	2.0	4.4	3.2
Form filtration rate (Kg/hr-sq m (including chemicals)	42.0	92.2	76.3	57.0	26.1	36.0
Cake solids content, weight percent	34.8	33.9	32.0	32.3	34.4	34.2
Total solids recovery, weight percent	99	95	98	96	99	98

(continued)

TABLE 37. (continued)

	RUN NUMBER					
	7	8	9	10	11	12
Feed total solids, g/ℓ	44.6	44.6	44.6	44.6	44.6	44.6
Conditioned feed total solids, g/ℓ	45.5	45.5	47.4	47.4	47.4	47.4
Lime dose, weight percent	10	10	20	20	20	20
Filter form time (θ_f), min.	0.75	1.3	2.0	1.0	0.75	1.3
Filter dry time (θ_d), min.	1.5	2.6	3.8	1.9	1.5	2.6
Form vacuum, mm Hg	530	530	510	530	510	510
Dry vacuum, mm Hg	510	510	480	480	480	510
Discharge code	G	G-E	G-E	G	F	G-E
Cake thickness, mm	1.2	1.6	1.6	1.6	1.2	1.6
Cake dry weight (W), Kg/sq m	0.51	0.57	0.65	0.54	0.51	0.77
θ_d/W, Kg-min/sq m	2.9	4.6	5.8	3.5	2.9	3.4
Form filtration rate (Kg/hr-sq m (including chemicals)	40.8	26.3	19.5	32.4	40.8	35.5
Cake solids content, weight percent	38.8	35.0	35.7	34.4	34.5	36.5
Total solids recovery, weight percent	97	99	99	98	95	98

(continued)

TABLE 37. (continued)

	RUN NUMBER					
	13	14	15	16	17	18
Feed total solids, g/ℓ	55.3	55.3	55.3	55.3	55.3	55.3
Conditioned feed total solids, g/ℓ	61.2	61.2	61.2	61.2	58.7	58.7
Lime dose, weight percent	19	19	19	19	10	10
Filter form time (θ_f), min.	1.1	0.41	0.59	0.36	1.1	0.75
Filter dry time (θ_d), min.	2.2	0.80	1.2	0.69	2.2	1.5
Form vacuum, mm Hg	430	510	510	510	510	510
Dry vacuum, mm Hg	360	460	460	480	480	460
Discharge code	E	E	E	G-F	G	G-F
Cake thickness, mm	4.0	2.0	3.2	2.4	2.4	1.6
Cake dry weight (W), Kg/sq m	1.4	0.88	1.0	0.77	0.95	0.78
θ_d/W, Kg-min/sq m	3.1	0.91	1.2	0.90	2.3	1.9
Form filtration rate (Kg/hr-sq m (including chemicals)	76.4	129	102	128	51.8	62.4
Cake solids content, weight percent	33.0	35.1	34.8	34.2	34.2	33.8
Total solids recovery, weight percent	99	99	99	98	99	99

(continued)

TABLE 37. (continued)

	RUN NUMBER				
	19	20	21	22	23
Feed total solids, g/ℓ	55.3	60.2	60.2	60.2	60.2
Conditioned feed total solids, g/ℓ	58.7	70.2	70.2	70.2	70.2
Lime dose, weight percent	10	26	26	26	26
Filter form time (θ_f), min.	2.0	2.0	0.59	0.41	0.32
Filter dry time (θ_d), min.	3.8	3.8	1.2	0.80	0.62
Form vacuum, mm Hg	510	330	380	380	460
Dry vacuum, mm Hg	430	280	330	360	380
Discharge code	E	E	E	G-E	G-F
Cake Thickness, mm	2.4	4.8	3.2	2.4	2.4
Cake dry weight (W), Kg/sq m	1.1	1.5	1.2	1.1	0.93
θ_d/W, Kg-min/sq m	3.5	2.5	1.0	0.73	.67
Form filtration rate (Kg/hr-sq m (including chemicals)	33.0	45.0	102	161	174
Cake solids content, weight percent	36.4	30.7	31.8	32.4	32.8
Total solids recovery, weight percent	99	99	96	96	96

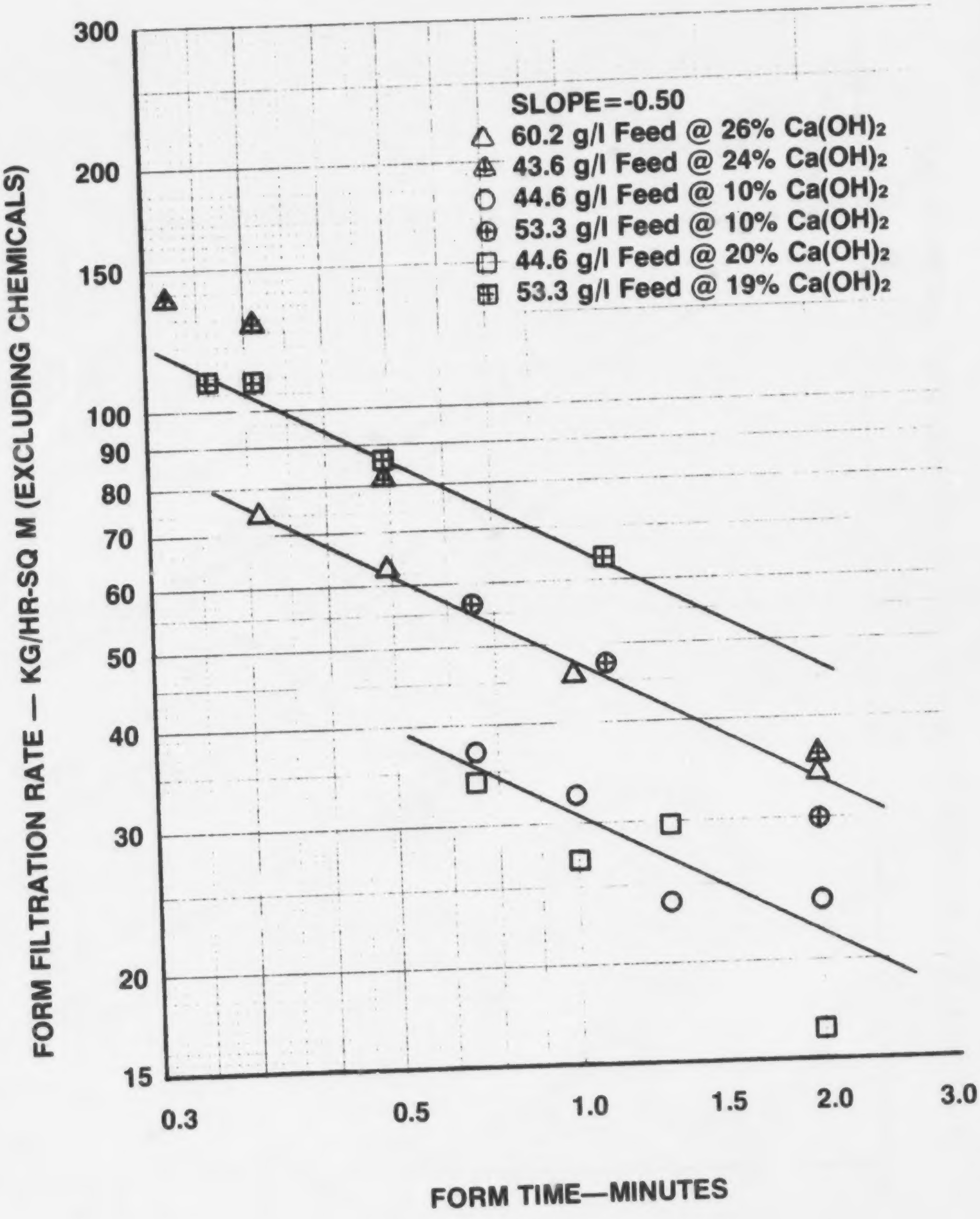

FIGURE 67. FORM FILTRATION RATE VS. FORM TIME FOR FERRIC-PRIMARY SLUDGE — PHASE 2. (High Ferric; with Polymer)

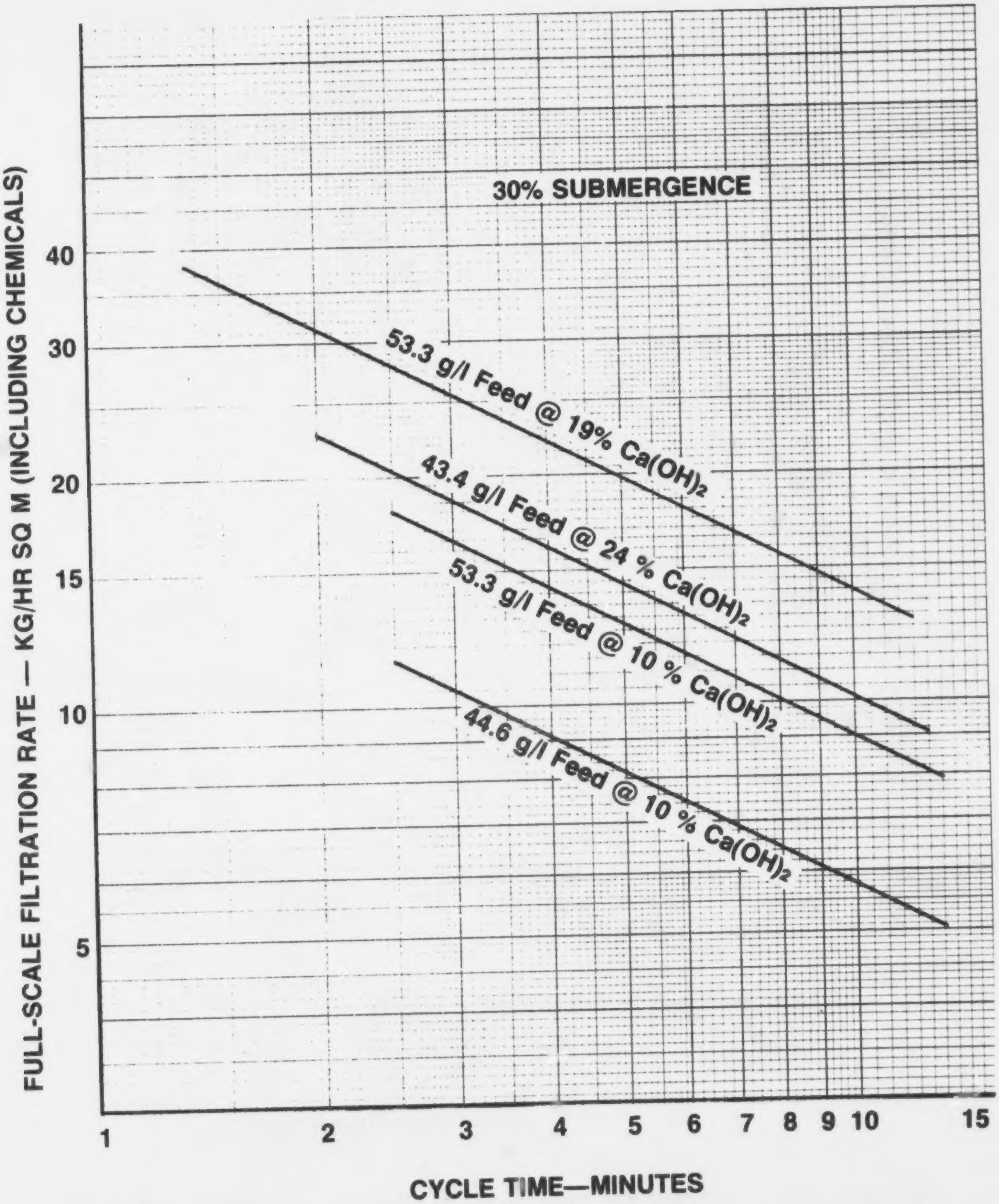

FIGURE 68. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR FERRIC-PRIMARY SLUDGE — PHASE 2.
(High Ferric; with Polymer)

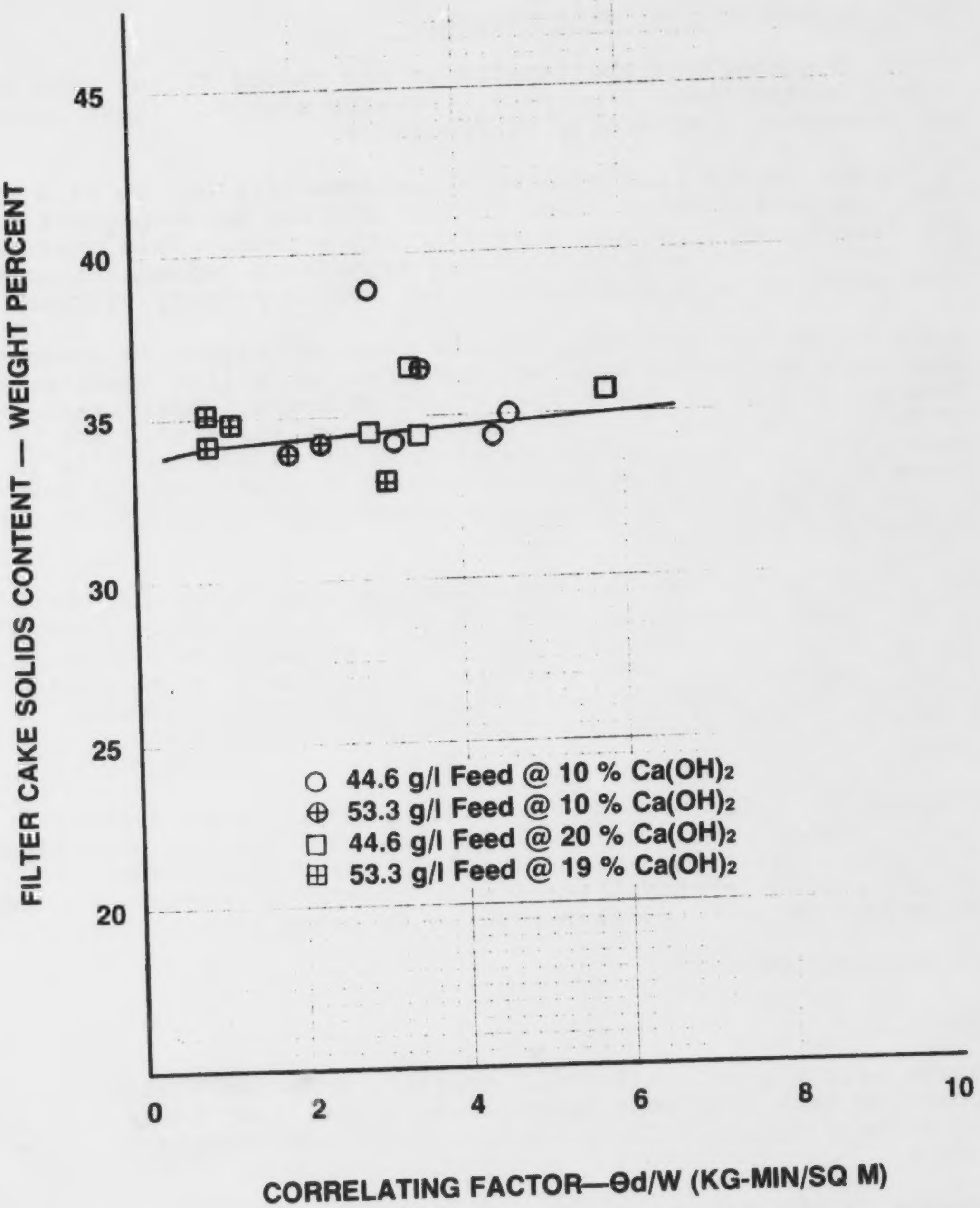

FIGURE 69. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR FERRIC-PRIMARY SLUDGE — PHASE 2.
(High Ferric; with Polymer)

Phase 3 (Low Ferric; with Polymer)

Table 38 summarizes the results of the vacuum filter runs performed during Phase 3 on ferric-primary sludge. These results are presented graphically on Figure 70.

As shown, feed solids concentration from 38.1 g/ℓ to 53.3 g/ℓ and lime conditioning doses from 16 percent to 56 percent did not significantly affect form filtration rate. This observation was inconsistent with much of the experience gained throughout this study on both alum-primary and ferric-primary sludges.

A review of the operating results shown on Figure 70 indicated that form filtration rate was maximized at a lime conditioning dose of 31 percent for a feed TS of 38.1 g/ℓ. Maximization of form filtration rate was also achieved for a 53.3 g/ℓ TS feed conditioning with lime at a dose of 16 percent. It is not known if the lime dose could have been further reduced for either of these feed conditions without any reduction of form filtration rate.

Figure 71 summarizes the two operating conditions discussed above. The information provided on Figure 71 and on Figure 72 may be used for full-scale predictive purposes as described in previous sections. Note that the results plotted on Figure 72 indicate that drier filter cakes were generally produced at the higher lime conditioning doses used during these filter runs.

Capture of vacuum filter feed sludge constituents evaluated during Phase 3 included volatile suspended solids, phosphorus and iron. Volatile suspended solids captures exceeded 99 weight percent whenever measured. Phosphorus capture averaged 92 weight percent and iron capture averaged 98 weight percent.

PRIMARY SLUDGE

Phase 4

Since only a relatively small quantity of primary sludge was produced each day, only bench-scale filter leaf tests were run on primary sludge. Table 39 summarizes the results of these tests.

Figure 73 shows a graphical summary of the bench-scale leaf test results for the primary sludge. As indicated, the slope of the form filtration rate - form time line was observed to be significantly greater than the theoretical -0.5. There are generally two reasons for a slope greater than -0.5.

TABLE 38

DATA SUMMARY OF VACUUM FILTRATION OF
FERRIC-PRIMARY SLUDGE FROM PHASE 3
(Low Ferric; with Polymer)

	RUN NUMBER						
	1	2	3	4	5	6	7
Feed total solids, g/ℓ	38.1	38.1	38.1	38.1	38.1	38.1	38.1
Conditioned feed total solids, g/ℓ	47.8	47.8	47.8	50.3	50.3	50.3	51.4
Lime dose, weight percent	31	31	31	45	45	45	56
Filter form time (θ_f), min.	2.0	1.3	1.0	2.0	1.3	1.0	2.0
Filter dry time (θ_d), min.	3.8	2.6	1.9	3.8	2.6	1.9	3.8
Form vacuum, mm Hg	510	510	510	510	530	510	510
Dry vacuum, mm Hg	460	460	460	480	480	480	480
Discharge code	F	P-F	P-ND	F	P-F	P-ND	F
Cake thickness, mm	1.2	0.8	0.8	1.2	0.8	0.8	1.2
Cake dry weight (W), Kg/sq m	0.54	.43	0.35	0.57	0.50	0.41	0.61
θ_d/W, Kg-min/sq m	7.0	6.0	5.4	6.7	5.2	4.6	6.2
Form filtration rate (Kg/hr-sq m (including chemicals)	16.2	19.8	21.0	17.1	23.1	24.6	18.3
Cake solids content, weight percent	34.0	33.3	34.3	37.4	36.3	36.2	37.3
Total solids recovery, weight percent	99	99	99	-	99	99	99

(continued)

TABLE 38. (continued)

	RUN NUMBER					
	8	9	10	11	12	13
Feed total solids, g/ℓ	38.1	53.3	53.3	53.3	53.3	53.3
Conditioned feed total solids, g/ℓ	51.4	62.8	62.8	62.8	65.4	65.4
Lime dose, weight percent	56	16	16	16	24	24
Filter form time (θ_f), min.	1.0	2.0	1.3	1.6	2.0	1.3
Filter dry time (θ_d), min.	1.9	3.8	2.6	3.1	3.8	2.6
Form vacuum, mm Hg	510	510	510	510	510	510
Dry vacuum, mm Hg	480	460	460	460	460	480
Discharge code	P	F-G	P-F	F	F-G	P
Cake thickness, mm	1.2	1.6	0.8	0.8	1.6	0.8
Cake dry weight (W), Kg/sq m	0.47	0.54	0.41	0.46	0.55	0.36
θ_d/W, Kg-min/sq m	4.0	7.0	6.3	6.7	6.9	3.6
Form filtration rate (Kg/hr-sq m (including chemicals)	28.2	16.2	18.9	17.3	16.5	16.6
Cake solids content, weight percent	36.4	34.8	33.6	33.4	34.8	37.9
Total solids recovery, weight percent	99	99	99	99	99	99

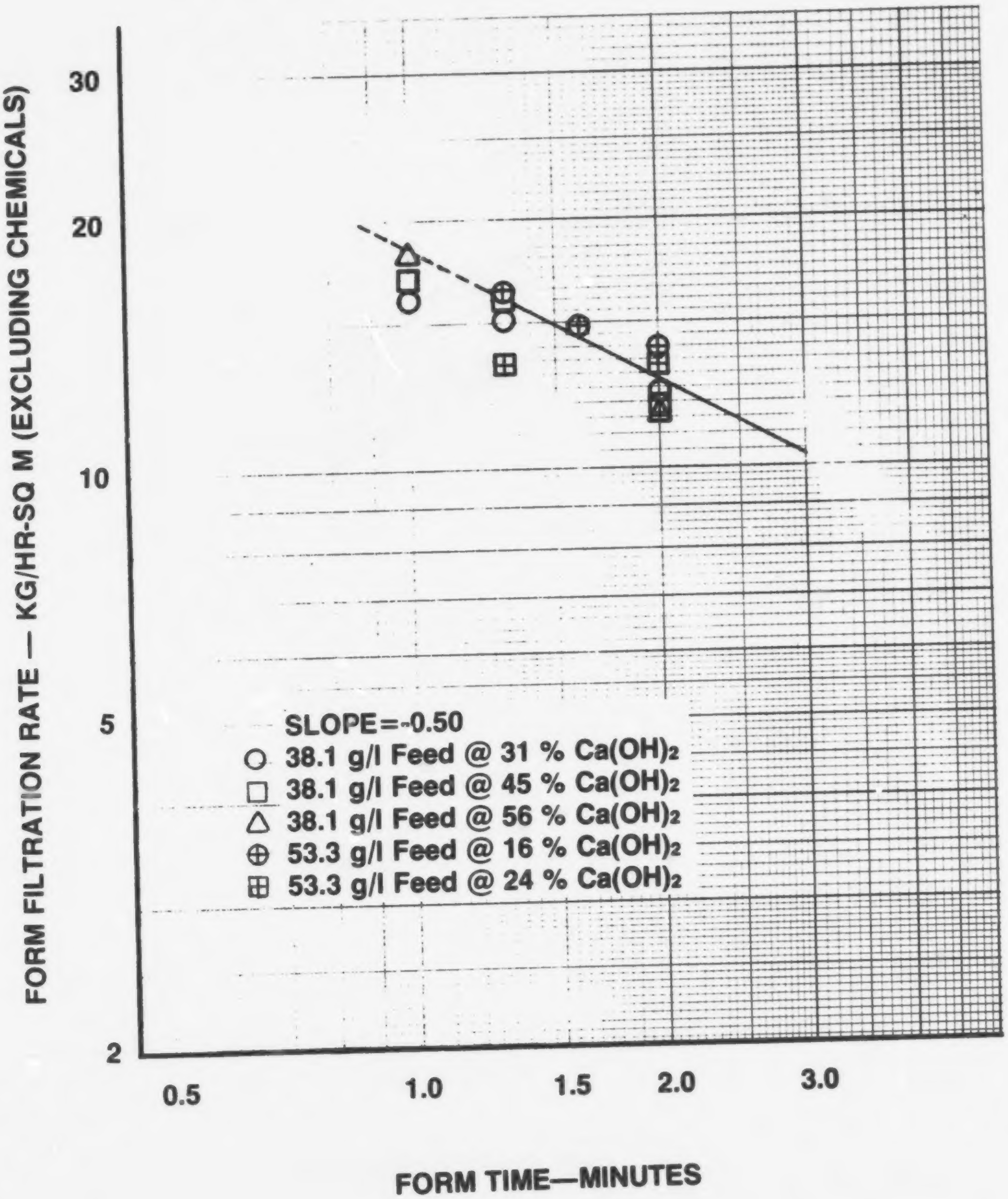

FIGURE 70. FORM FILTRATION RATE VS. FORM TIME FOR FERRIC-PRIMARY SLUDGE — PHASE 3.
(Low Ferric; with Polymer)

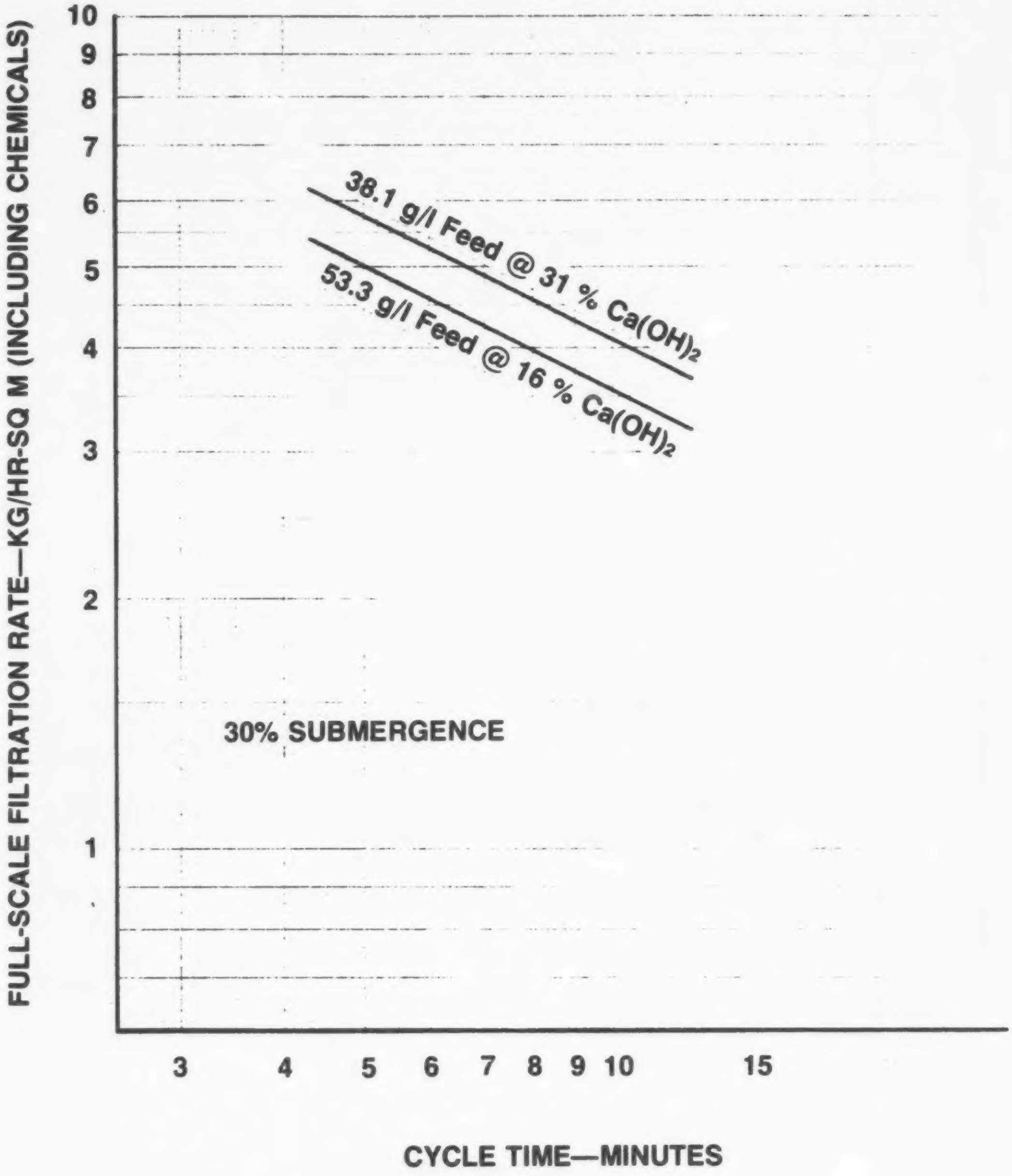

FIGURE 71. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR FERRIC- PRIMARY SLUDGE — PHASE 3.
(Low Ferric; with Polymer)

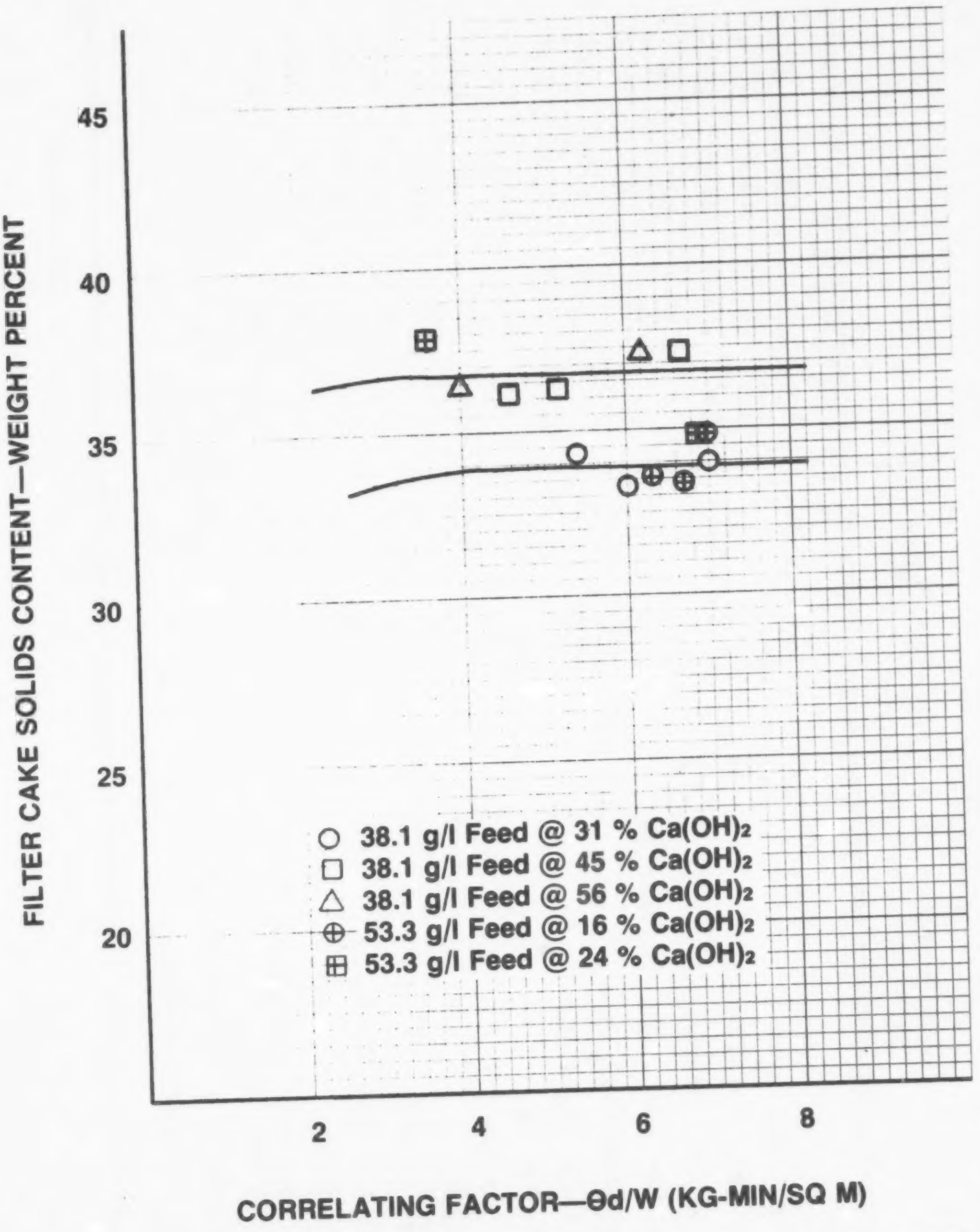

FIGURE 72. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR FERRIC-PRIMARY SLUDGE — PHASE 3.
(Low Ferric; with Polymer)

TABLE 39

DATA SUMMARY OF VACUUM FILTRATION LEAF TESTS OF PRIMARY SLUDGE FROM PHASE 4

	RUN NUMBER						
	1	2	3	4	5	6	7
Feed total solids, g/ℓ	35.2	35.2	35.2	35.2	35.2	40.6	40.6
Conditioned feed total solids, g/ℓ	39.2	39.2	39.2	39.2	39.2	47.0	47.0
Lime dose, weight percent	12	12	12	12	12	16	16
Ferric chloride dose, weight percent	3	3	3	3	3	4	4
Filter form time (θ_f), min.	1.0	0.5	2.0	4.0	2.0	1.0	0.5
Filter dry time (θ_d), min.	1.0	2.0	1.0	2.0	4.0	1.0	2.0
Form vacuum, mm Hg	510	510	510	510	510	510	510
Dry vacuum, mm Hg	510	510	510	510	510	510	510
Discharge code	G	E	F	F	E	G	F
Cake thickness, mm	5.6	4.8	7.9	12.7	7.1	3.6	2.4
Cake dry weight (W), Kg/sq m	1.2	0.86	1.3	1.7	1.3	0.87	0.52
θ_d/W, Kg-min/sq m	0.83	2.3	0.77	1.2	3.1	1.2	3.8
Form filtration rate (Kg/hr-sq m (including chemicals)	72.0	103	39.0	25.5	39.0	52.2	62.4
Cake solids content, weight percent	17.5	24.0	12.8	11.4	20.8	16.7	14.0
Total solids recovery, weight percent	95	95	95	95	95	92	92

(continued)

TABLE 39. (continued)

	RUN NUMBER							
	8	9	10	11	12	13	14	15
Feed total solids, g/ℓ	40.6	40.6	40.6	41.3	41.3	41.3	41.3	41.3
Conditioned feed total solids, g/ℓ	47.0	47.0	47.0	49.6	49.6	49.6	49.6	49.6
Lime dose, weight percent	16	16	16	20	20	20	20	20
Ferric chloride dose, wt. %	4	4	4	5	5	5	5	5
Filter form time (θ_f), min.	2.0	4.0	2.0	1.0	0.5	2.0	4.0	2.0
Filter dry time (θ_d), min.	1.0	2.0	4.0	1.0	2.0	1.0	2.0	4.0
Form vacuum, mm Hg	510	510	510	510	510	510	510	510
Dry vacuum, mm Hg	510	510	510	510	510	510	510	510
Discharge code	P	P	P-F	E	E	E	E	E
Cake thickness, mm	3.2	3.2	2.4	7.1	5.6	7.9	6.4	4.8
Cake dry weight (W), Kg/sq m	0.52	0.53	0.54	1.7	1.3	1.8	1.7	1.2
θ_d/W, Kg-min/sq m	1.9	3.8	7.4	.59	1.5	0.56	1.2	3.3
Form filtration rate (Kg/hr-sq m (including chemicals)	15.6	8.0	16.2	102	156	54.0	25.5	36.0
Cake solids content, wt. %	9.5	9.8	13.0	23.1	29.0	22.7	25.6	29.9
Total solids recovery, wt. %	92	92	92	98	98	98	98	98

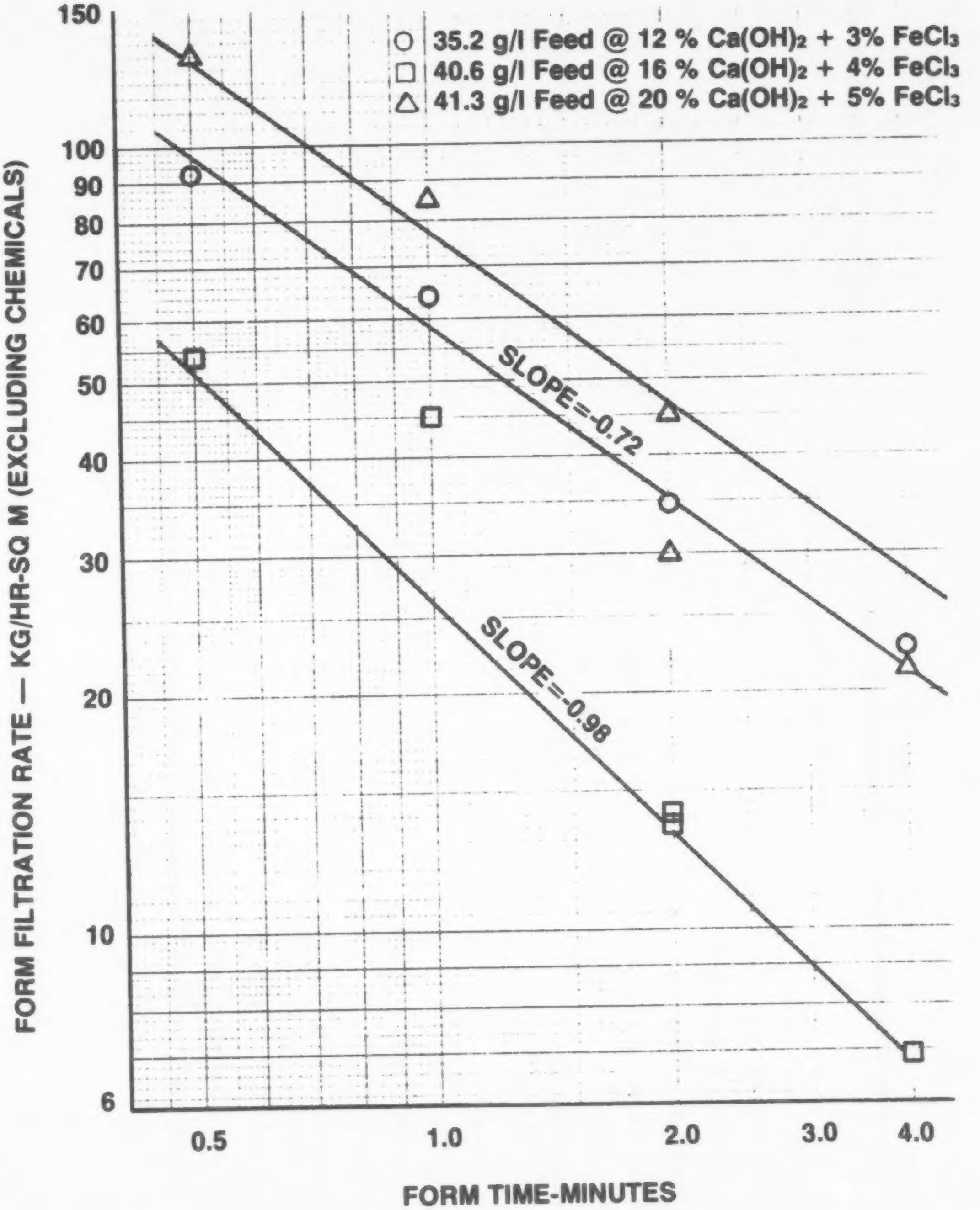

FIGURE 73.— FORM FILTRATION RATE VS. FORM TIME FOR PRIMARY SLUDGE — PHASE 4.

One suggests that sludge constituents of a significantly smaller particle size (fines) than the majority of the feed sludge tend to migrate into the pores of the filter cake during filtration. The net result is a less porous filter cake with relatively poorer filtration characteristics than if no particle migration occurred. The degree of particle migration and the magnitude of its effect on cake porosity and cake formation characteristics is determined by the specific particle size distribution of the feed sludge dry solids. With fine particle migration and longer cake form times, less filter cake is formed due to the less porous cake. The result would be, as observed, lower form filtration rates than expected at the longer cake form times.

The second explanation for a slope greater than -0.5 involves cake sloughing during formed cake removal from the feed slurry vat. The net effect of this, as observed, is lower form filtration rates than expected at the longer cake form times.

Observations made during the filter leaf tests indicate that cake sloughing was responsible for the slopes which were calculated. The filter cake appeared to form very quickly with very little additional cake formed at longer form times. This is best illustrated by the results plotted on Figure 73 which indicated a slope of -0.98. The maximum possible negative slope is -1.0. Observation of a -1.0 slope indicates that no additional cake is formed over that formed at the lowest form time evaluated. Evidently, this extreme situation was occurring during the leaf tests which resulted in the -0.98 slope shown on Figure 73.

A review of Figure 73 also indicates anomalous results for the data with a -0.98 slope. The form filtration rates observed at a lime dose of 16 percent were lower than those observed for the tests conducted at lime doses of 12 or 16 percent. Because of the observed slope of this data and because of its inconsistency with the remainder of the leaf test data, the results of the tests conducted at a 16 percent lime dose are considered suspect, and were eliminated from the predicted full-scale performance curves shown on Figures 74 and 75.

The scale-up of 0.8 from leaf test results to full-scale filter performance is based on Eimco-PMD experience. Note on Figure 75 that significantly drier cakes were produced with a lime conditioning dose of 20 percent as compared to 12 percent.

DISCUSSION

The method chosen to rationalize vacuum filter performance measured during the various phases of this study was based on the chemical constituent content of the chemical-primary sludges produced.

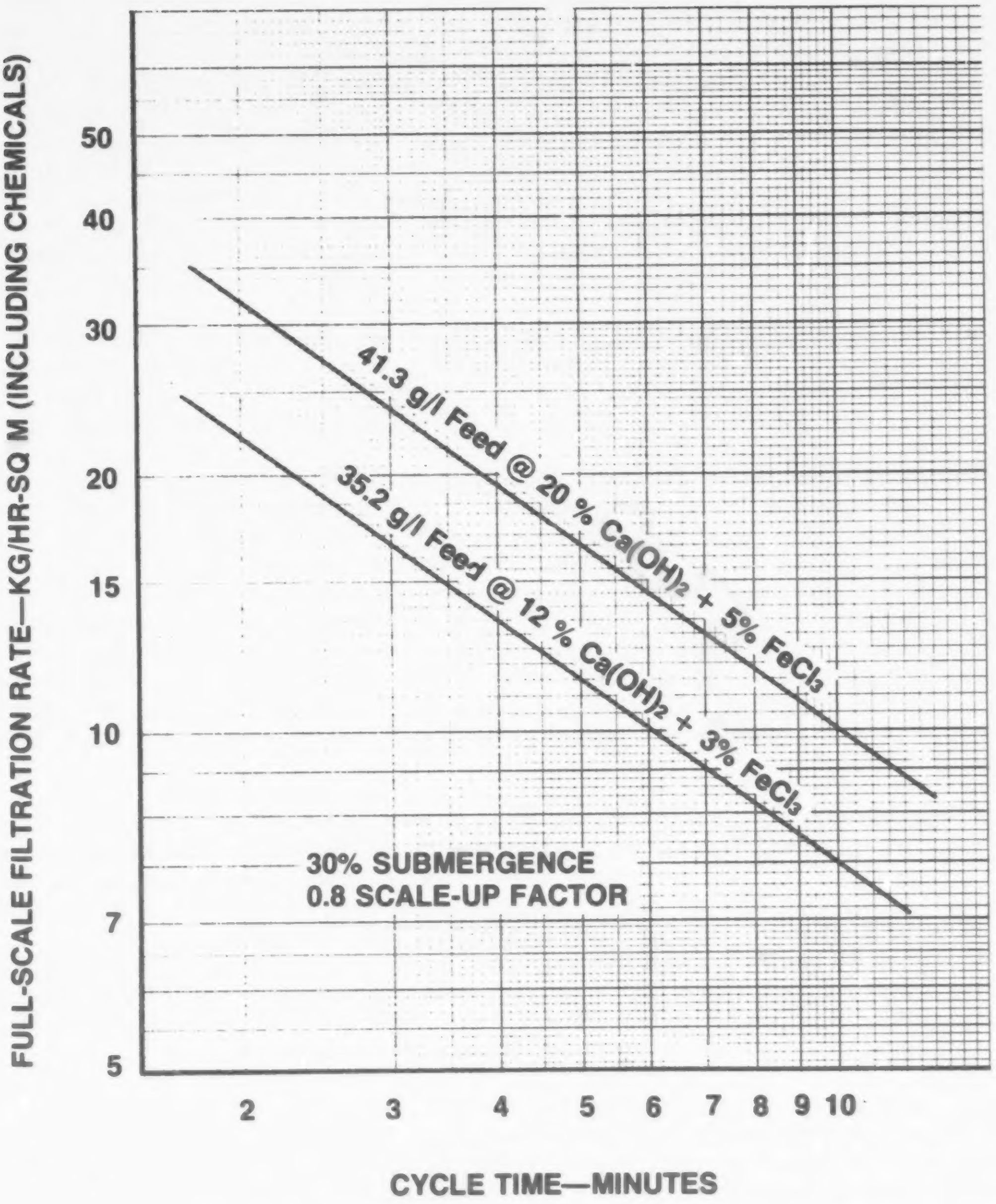

FIGURE 74. FULL-SCALE FILTRATION RATE VS. CYCLE TIME FOR PRIMARY SLUDGE — PHASE 4.

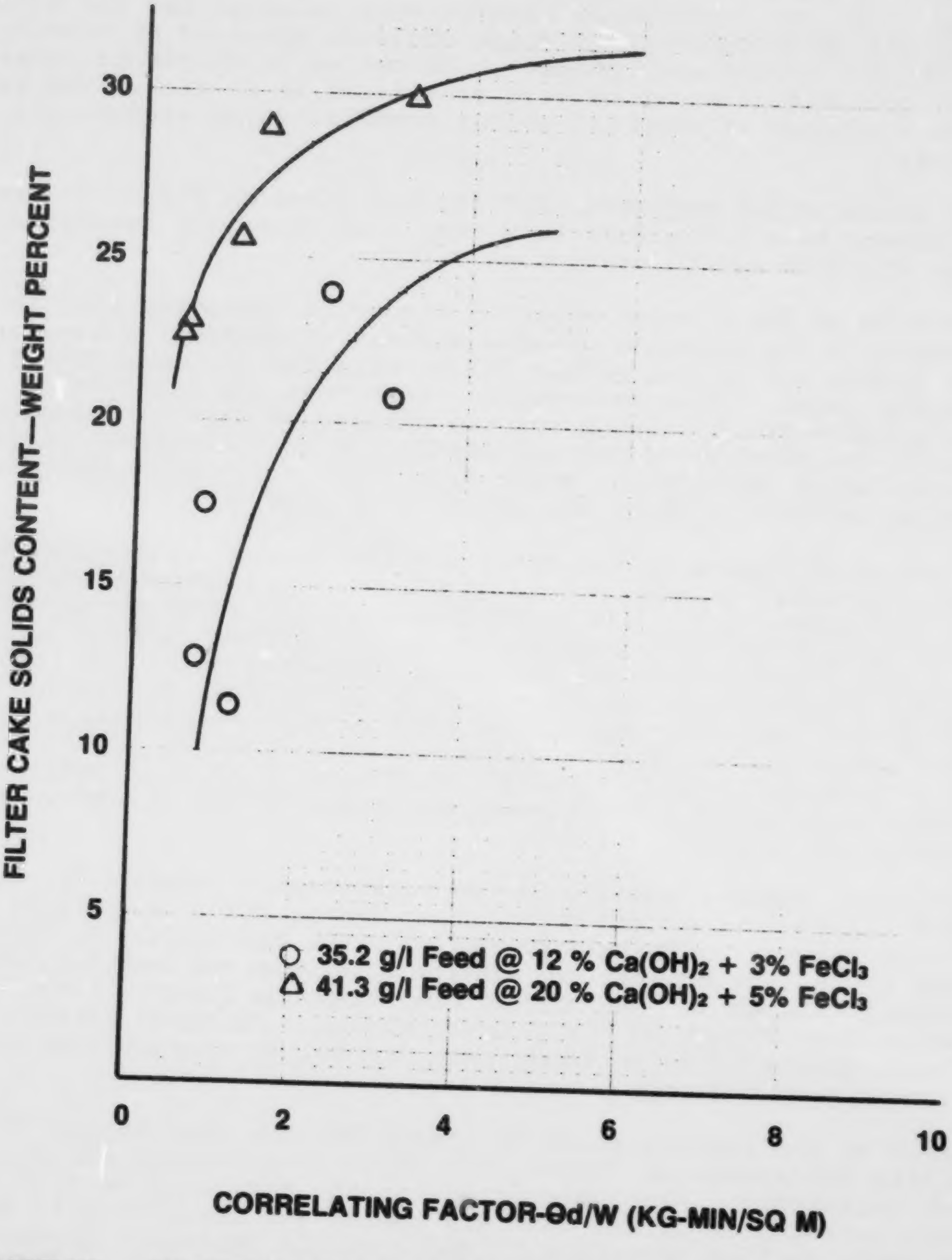

FIGURE 75.— FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR PRIMARY SLUDGE — PHASE 4.

Figure 76 shows the relationship between maximum form filtration rate since the performance results were selected for the maximum feed solids concentrations which could be obtained by gravity thickening during each phase. Differences in thickener underflow solids concentration were considered to be due to the relative fractions of chemical solids comprising the sludges produced.

For illustrative purposes, the results shown on Figure 76 are pertinent to a 1.5 minute form time, and to a lime conditioning dose of 25-30 weight percent as $Ca(OH)_2$.

A review of the results shown on Figure 76 indicates that increases in the fraction of alum associated chemical solids in the sludge fed to the vacuum filter resulted in lower form filtration rates. This observation means that the addition of filter alum to raw sewage for the purpose of improving suspended solids and phosphorus removal results in an alum-primary type sludge which dewaters at lower filtration rates than the primary sludge produced without the addition of alum.

As shown on Figure 76 the deterioration in form filtration rate with increases in the percent alum associated chemical solids in the sludge is extreme. The curve was not drawn through the zero percent chemical solids data point (primary sludge from Phase 4) since filtration data were collected only at a feed sludge concentration of 41.3 g/ℓ. A higher form filtration rate would have been observed if the feed sludge concentration had been closer to the gravity thickener underflow (70 to 100 g/ℓ TS). Nevertheless, the results on Figure 76 show a dramatic effect of alum associated chemical solids in the sludge on form filtration rate.

Figure 77 shows a summary of the relationship between filter cake solids content and correlating factor for alum-primary sludge dewatering for all phases. Little difference in filter cake dry solids content was observed between the results for Phase 1 through 4. The dotted line drawn on Figure 77 was transferred from Figure 74 for primary sludge. As shown primary sludge dewatered to a higher cake dry solids content than the alum-primary from any phase.

Based on the results shown on Figure 76, the observation of little difference in filter cake dry solids between the phases was unexpected. Also unexpected was the fact that no significant increase in filter cake dry solids was observed with an increase in lime dose from 15 to 37 weight percent. However, it can be noted that the addition of filter alum to raw sewage resulted in filter cakes with higher moisture contents.

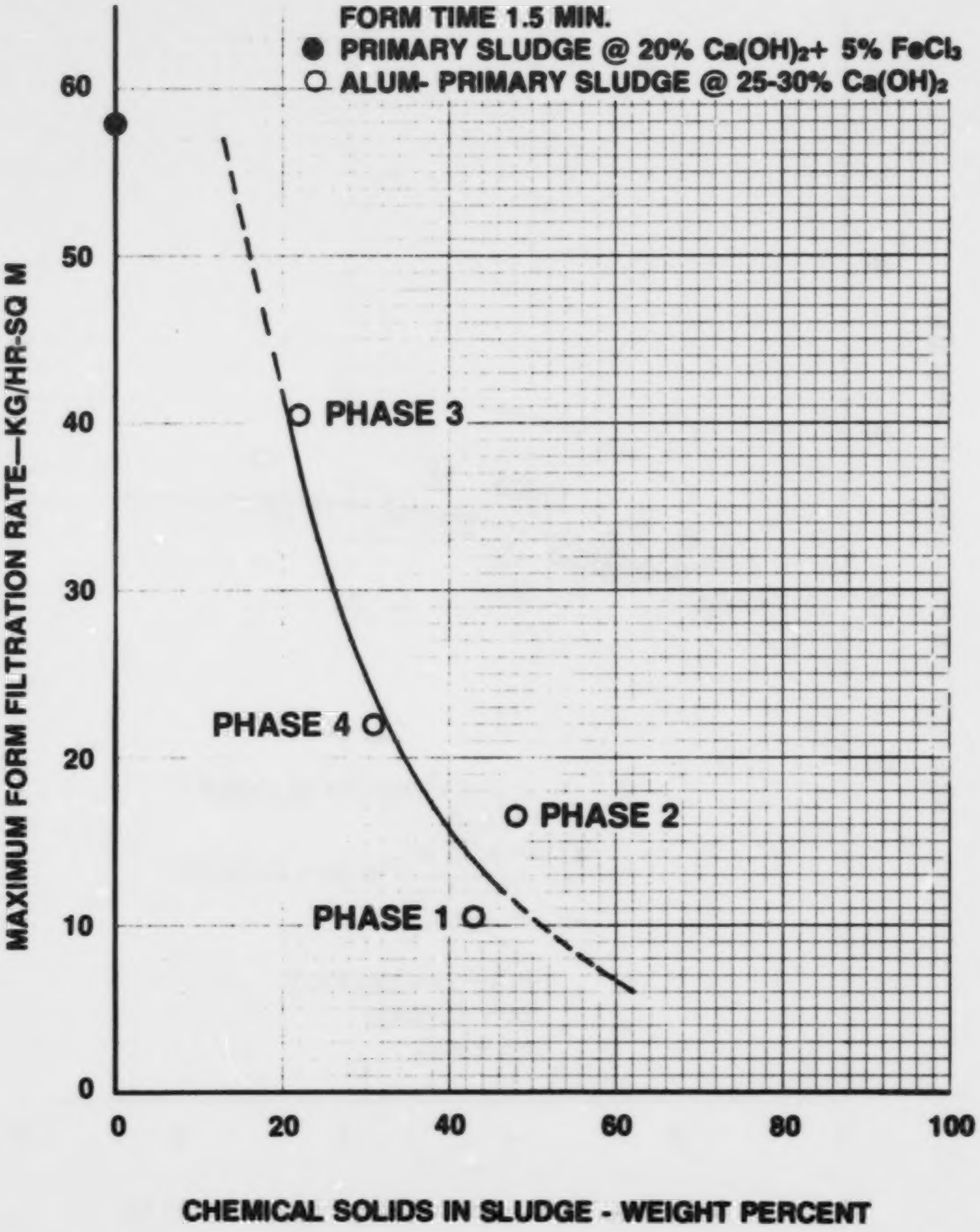

FIGURE 76. EFFECT OF CHEMICAL SOLIDS CONTENT OF ALUM-PRIMARY SLUDGE ON FORM FILTRATION RATE.

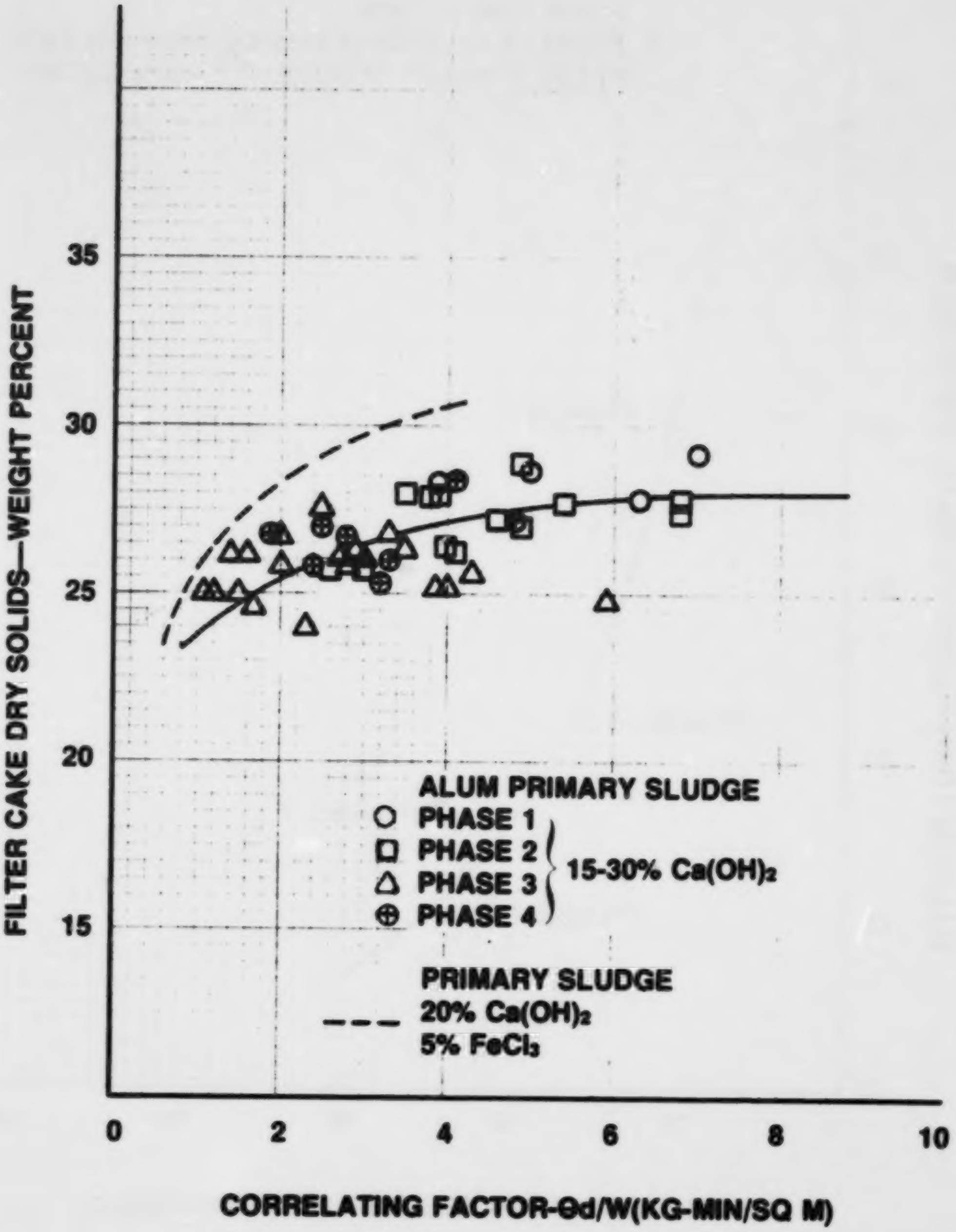

FIGURE 77. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR ALL ALUM-PRIMARY SLUDGE PHASES.

A similar analysis was developed for ferric-primary sludge, vacuum filtration dewatering. Figures 78 and 79 show the results of this analysis.

An obvious relationship between maximum form filtration rate and the fraction of chemical solids in the sludge was not documented. Figure 78 shows that extremely erratic results were obtained. The results from Phases 1 and 3 indicated poor filtration performance while excellent filtration was achieved during Phase 2.

Difficulty in vacuum filtration dewatering of ferric-primary sludge was experienced during a previous study.[2] In this study, raw wastewater was contacted with ferric chloride in a Reactor-Clarifier operated in a solids contact mode. The ferric-primary sludge proved to be nearly impossible to filter on a three foot diameter by three foot face vacuum filter. Acceptable cake discharge was not achieved even at lime conditioning doses of up to 50 weight percent as $Ca(OH)_2$.

The filtration problem experienced during this former study was tentatively traced to the presence of ferrous sulfide in the sludge. It was hypothesized that anaerobic conditions within the sludge blanket of the Reactor-Clarifier caused the reduction of the ferric to ferrous iron with the formation of ferrous sulfide.

The approach followed during the course of this work was to maintain as low a sludge solids retention time as possible within the flocculating clarifier. This was done in order to maintain as "fresh" a sludge as possible with the goal of preventing ferric reduction.

A review of the results shown on Figure 78 indicates that the ferric-primary sludge produced during Phase 2 exhibited filtration characteristics greatly superior to that produced in Phases 1 or 3. This observation was inconsistent with the hypothesis that increasing fractions of chemical solids result in poorer filtration characteristics. No definitive conclusion may be drawn from the results shown on Figure 78 as to the effect of chemical solids in the ferric-primary sludge on filtration characteristics.

It seems apparent, therefore, that the ferric-primary sludges produced during this study were exposed to undocumented conditions which drastically affected their thickening and dewatering characteristics. It was hypothesized that the ferric-primary sludges produced during Phases 1 and 3 may have been retained in the chemical treatment unit for too long a period of time, and that this may have resulted in ferric iron reduction.

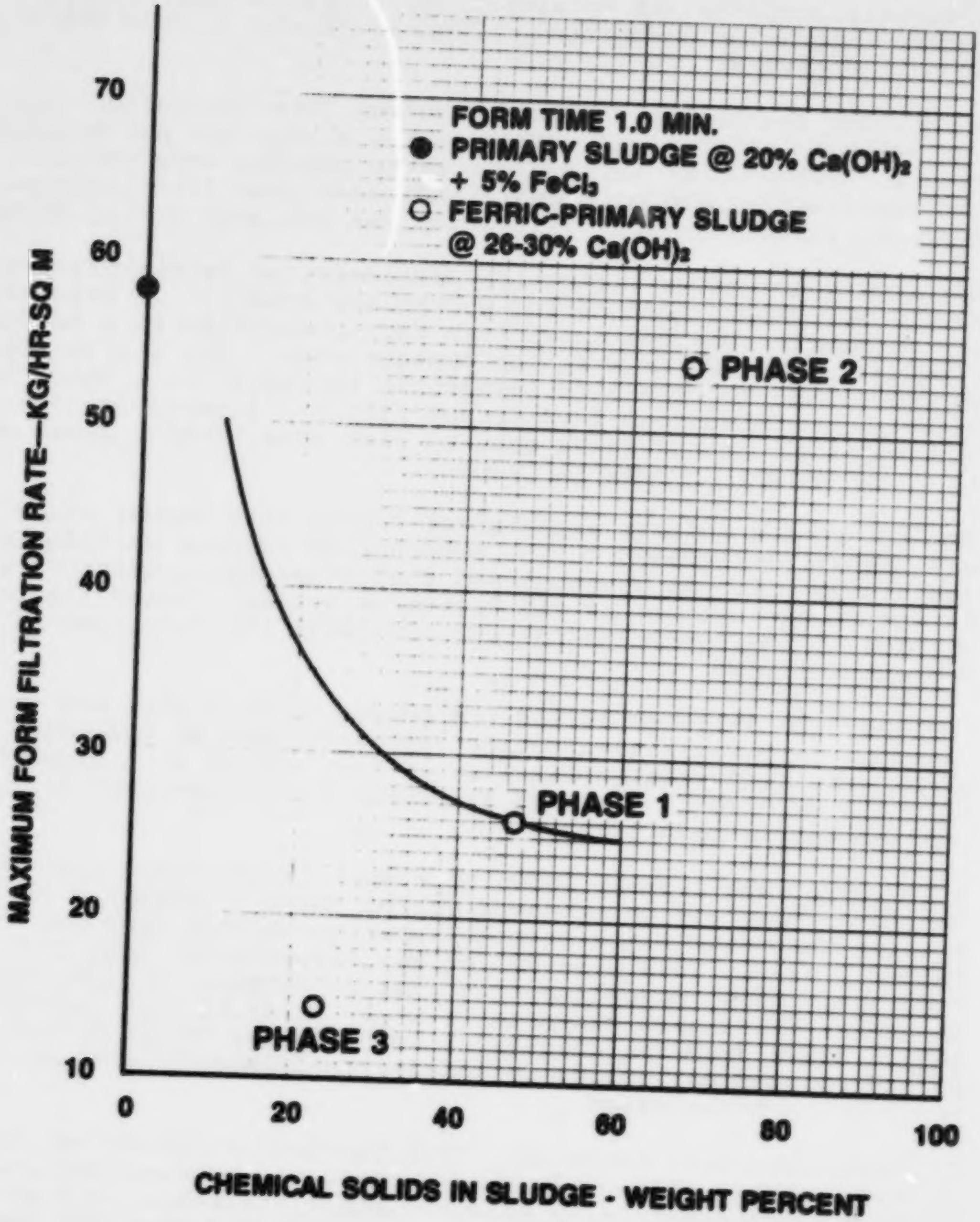

FIGURE 78. EFFECT OF CHEMICAL SOLIDS CONTENT OF FERRIC-PRIMARY SLUDGE ON FORM FILTRATION RATE.

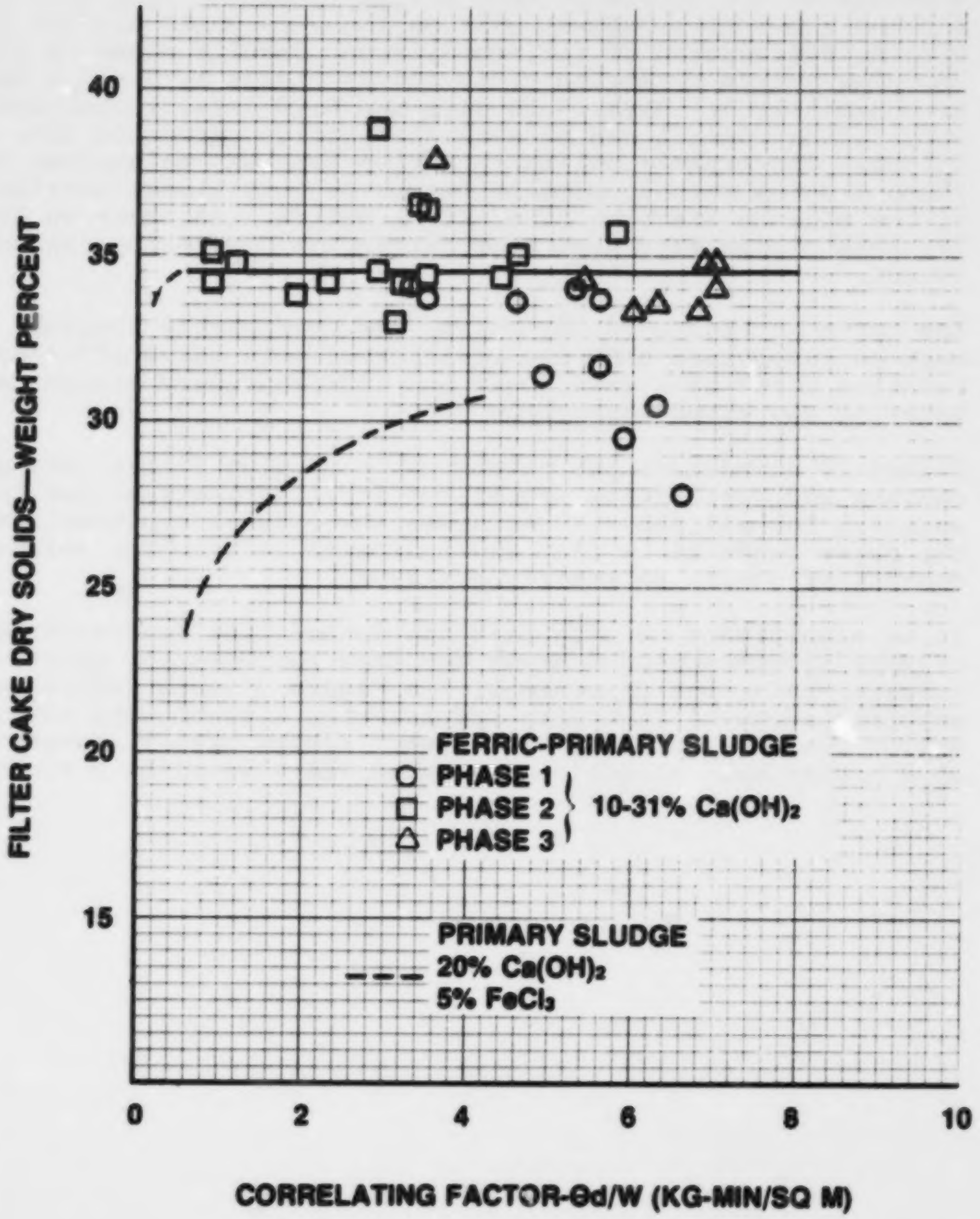

FIGURE 79. FILTER CAKE SOLIDS CONTENT VS. CORRELATING FACTOR FOR ALL FERRIC-PRIMARY SLUDGE PHASES.

A review of the ferric-primary solids retention time maintained for the chemical treatment unit during each phase did not provide an explanation of the inconsistent results shown on Figure 78. The solids retention times during Phases 1, 2 and 3 were approximately 4.5 days, 0.50 days and 0.25 days, respectively. Only during Phase 1 was an excessive solids retention time maintained. The minimum solids retention time was maintained during Phase 3 which corresponded to ferric-primary sludge consisting of the minimum fraction of chemical solids. As shown on Figure 78, this sludge exhibited poor filtration dewatering characteristics.

The curve on Figure 78, as shown, was arbitrarily located. It must be recognized that the location of this curve affected the relative costs of alum-primary and ferric-primary sludge dewatering described in Section VII.

Figure 79 summarizes the relationship between filter cake solids content and correlating factor for ferric-primary sludge dewatering for all phases. Although the results are scattered, the curve tends to be flat and indicates that a cake solids content of 33-35% is readily achieved.

It is significant to note that the dotted line representing primary sludge appeared below the data representing ferric-primary sludge solids content. On Figure 79 this dotted line appeared above all the data representing alum-primary sludge solids content; thus, ferric-primary sludge clearly dewatered to a higher cake solids content than did alum-primary sludge.

CENTRIFUGAL DEWATERING

Alum-Primary Sludge

Phase 1 (High Alum; no Polymer)

Performance of the centrifuge on alum-primary sludge generated during Phase 1 is shown in Table 40 and on Figures 80-81.

For a feed concentration of 23-24 g/ℓ, 0.70-0.90 percent by weight of dry sludge solids of anionic polymer (Dow AP-30) was required to achieve total solids capture greater than 80 percent. Highest solids capture using this polymer dosage was 98 percent.

Solids capture of only about 20 percent was achieved during operation without polymer. Highest attainable cake dry solids were 30-31 percent at very low (20 percent) solids capture. Solids captures in excess of 40 percent resulted in cake dry solids contents of 15-16 percent.

Phase 2 (High Alum; with Polymer)

Table 41 and Figures 82-83 illustrate the results of all centrifugal dewatering runs performed on alum-primary sludge during Phase 2.

For a feed concentration of 27 g/ℓ, 0.31-0.55 percent polymer by weight was required to achieve solids capture greater than 85 percent. Highest solids capture using this polymer dosage was 94-99 percent. Solids capture of only 20-30 percent was achieved during operation without polymer.

As shown in Figure 82, an anomaly was observed. Using a polymer dosage of 0.22-0.28 percent resulted in a decrease in solids capture as the hydraulic loading rate to the machine decreased. No reason can be offered for this inconsistency.

Cake solids relationships were essentially the same as those in Phase 1. Highest attainable cake dry solids were 30-31 percent at low (20 percent) solids captures, while solids captures greater than 40 percent resulted in cake solids of 15-16 percent.

Phase 3 (Low Alum; with Polymer)

Results of all centrifugal dewatering runs performed on alum-primary sludge generated during Phase 3 are illustrated in Table 42 and on Figures 84-85.

TABLE 40

DATA SUMMARY OF CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE FROM PHASE 1

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
1	20	22.5	3.14	0	30.1	21.0
2	20	22.5	4.04	0	28.4	20.6
3	20	22.5	7.28	0	27.1	19.3
4	20	22.5	3.23	0.24	19.4	33.5
5	20	22.5	3.91	0.26	13.9	39.9
6	20	22.5	7.58	0.20	18.1	23.4
7	20	22.5	12.8	0.19	16.3	51.8
8	20	22.5	3.37	0.50	15.6	58.6
9	20	22.5	7.58	0.38	14.7	48.7
10	20	22.5	12.8	0.33	15.2	42.2
11	20	22.5	4.04	0.75	16.7	97.7
12	20	22.5	7.24	0.90	15.9	96.1
13	12	22.5	4.18	0	23.1	22.8
14	25	22.5	3.64	0	31.5	21.9
15	20	22.5	12.8	0.70	16.0	82.3
16	12	23.9	3.97	0	28.9	19.9
17	20	23.9	3.50	0	29.8	21.2
18	20	23.9	4.18	0	29.9	20.7
19	20	23.9	8.35	0	27.9	18.2
20	25	23.9	4.38	0	30.8	20.7

*Dow AP-30

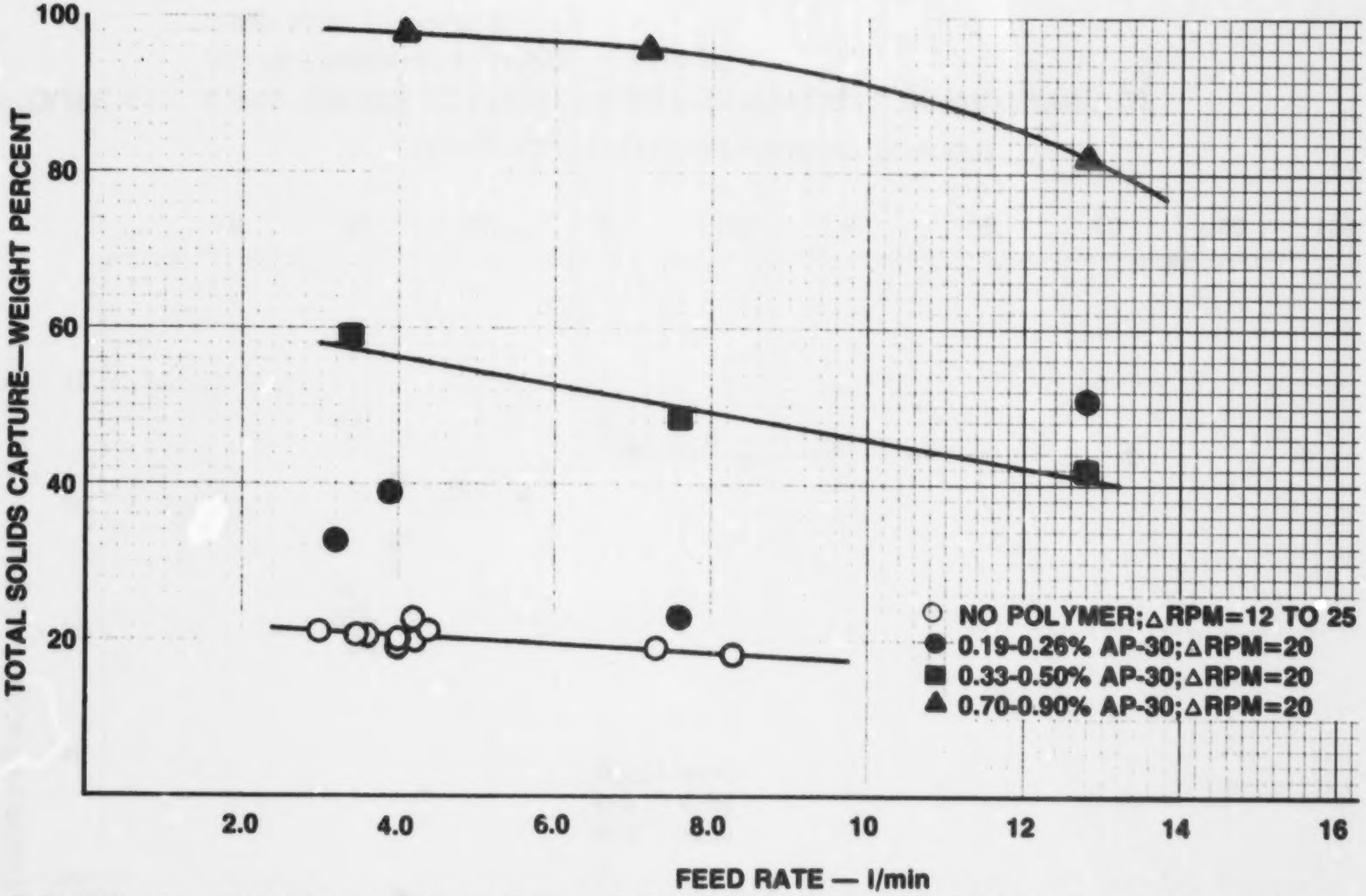

FIGURE 80. SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE — PHASE 1.
(High Alum; no Polymer)

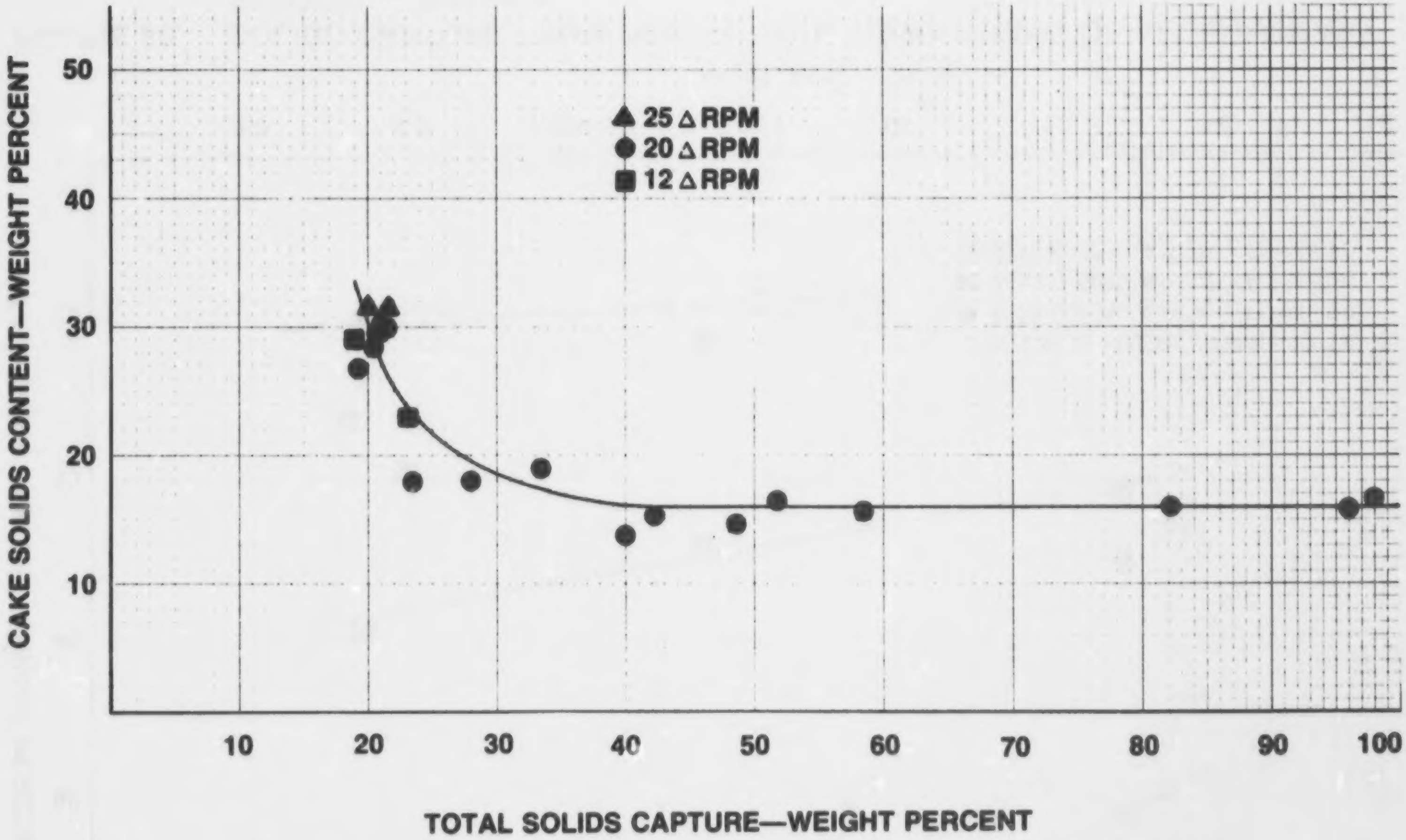

FIGURE 81. CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE — PHASE 1.
(High Alum; no Polymer)

TABLE 41

DATA SUMMARY OF CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE FROM PHASE 2

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
1	10	27.2	3.55	0	25.4	24.5
2	10	27.2	6.42	0	28.4	21.9
3	25	27.2	3.71	0	29.9	22.6
4	25	27.2	6.33	0	29.3	21.4
5	20	27.2	3.62	0	30.5	19.4
6	20	27.2	6.33	0	27.4	20.8
7	20	27.2	11.8	0	29.6	20.3
8	20	27.2	15.2	0	28.0	19.2
9	20	27.2	3.77	0.13	16.0	32.6
10	20	27.2	6.20	0.13	16.7	32.4
11	20	27.2	8.76	0.13	16.6	30.1
12	20	27.2	11.8	0.14	17.1	28.4
13	20	27.2	1.35	0.22	16.4	39.2
14	20	27.2	6.29	0.27	15.2	36.8
15	20	27.2	8.59	0.27	14.4	72.0
16	20	27.2	12.0	0.23	16.4	78.9
17	20	27.2	6.23	0.47	15.0	93.0
18	20	27.2	8.76	0.38	15.2	58.9
19	20	27.2	12.1	0.35	13.7	85.0
20	20	27.2	15.2	0.36	14.9	93.3
21	20	27.2	6.39	0.53	14.8	93.3
22	20	27.2	8.93	0.55	15.0	93.9

(continued)

*Dow AP-30

TABLE 41. (continued)

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
22	20	27.2	8.93	0.55	15.0	93.9
23	20	27.2	11.9	0.48	15.1	93.0
24	20	27.2	16.2	0.40	15.9	99.4
25	10	26.6	3.64	0	26.4	22.9
26	10	26.6	5.73	0	28.5	22.7
27	25	26.6	3.71	0	24.2	25.0
28	25	26.6	6.51	0	29.0	20.7
29	20	26.6	3.75	0	25.1	26.6
30	20	26.6	6.06	0	30.1	23.4
31	20	26.6	11.6	0	29.6	20.3
32	20	26.6	14.9	0	26.8	18.8
33	20	26.6	3.97	0.14	16.3	31.5
32	20	26.6	6.06	0.12	18.4	33.4
35	20	26.6	8.53	0.09	17.1	31.3
36	20	26.6	11.6	0.08	20.7	34.6
37	20	26.6	4.04	0.19	15.8	43.2
38	20	26.6	5.66	0.28	14.8	53.8
39	20	26.6	8.09	0.25	17.5	85.3
40	20	26.6	11.6	0.22	19.3	72.9
41	20	26.6	5.93	0.38	16.7	93.3
42	20	26.6	8.22	0.33	15.0	99.0
43	20	26.6	11.7	0.31	15.3	84.9
44	20	26.6	14.8	0.28	15.2	84.6
45	20	26.6	5.95	0.47	17.9	99.0
46	20	26.6	8.59	0.46	17.2	95.8
47	20	26.6	11.8	0.43	15.4	97.2
48	20	26.6	14.6	0.43	16.7	99.4

*Dow AP-30

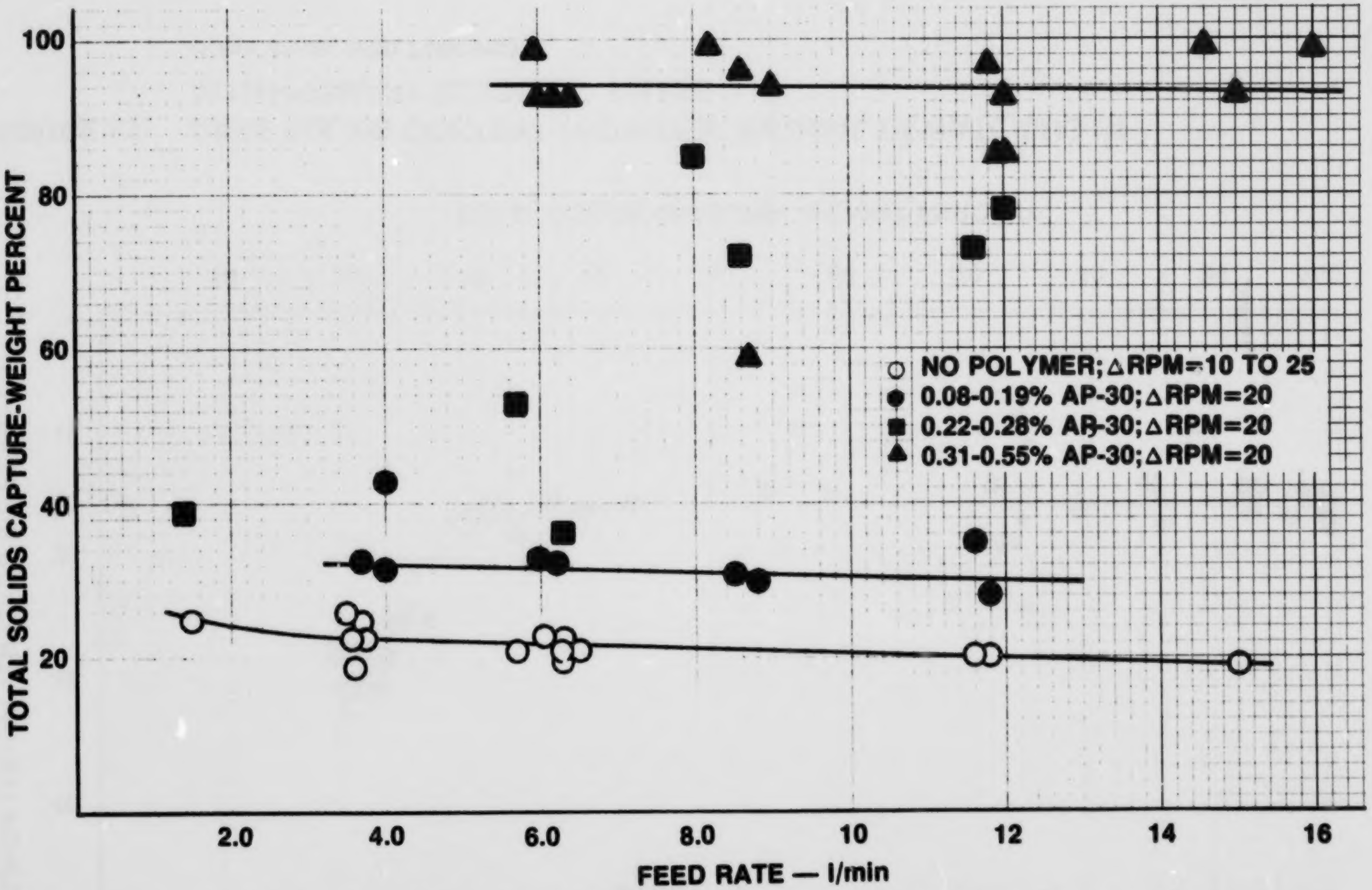

FIGURE 82. SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE — PHASE 2.
(High Alum; with Polymer)

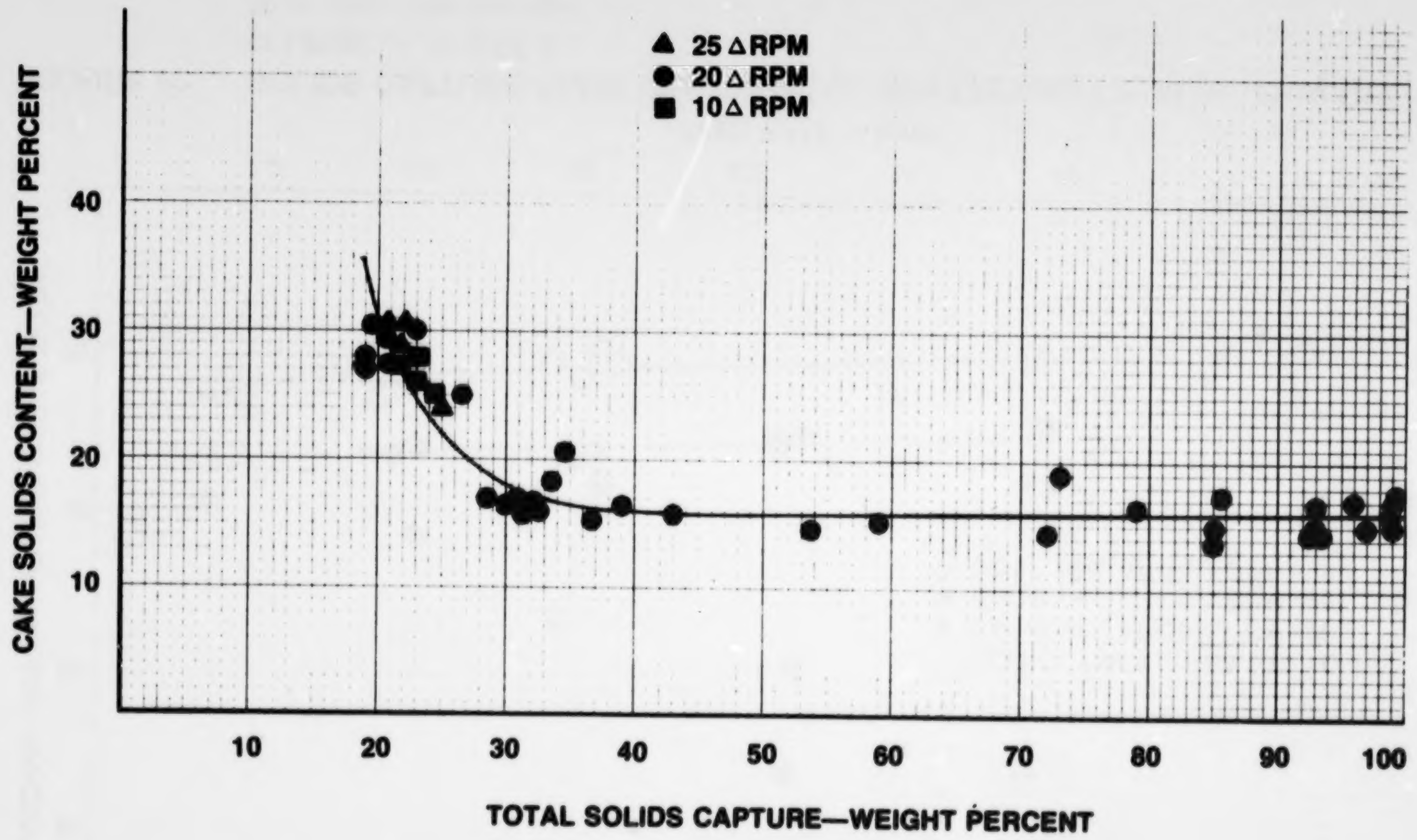

FIGURE 83. CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE — PHASE 2.
(High Alum; with Polymer)

For a feed concentration of 45-47 g/ℓ and using a polymer dosage of 0.21-0.25 percent, 90-93 percent solids capture was attained, while a polymer dosage of 0.30-0.40 percent by weight resulted in solids captures of 96-99 percent.

Using no polymer resulted in solids captures as high as 35-40 percent. Since a higher solids capture with no polymer was achieved (compared to 20 percent during Phases 1 and 2), maximum cake solids were slightly lower, at 27-28 percent. Cake solids of 17-19 percent were achieved consistently at solids captures above 50 percent.

FERRIC-PRIMARY SLUDGE

Phase 1 (High Ferric; no Polymer)

Results of all centrifugal dewatering runs performed on ferric-primary sludge generated during Phase 1 are shown in Table 43 and on Figures 86-87.

For a feed concentration of 34 g/ℓ, a polymer dose of 0.25-0.60 percent by weight was required to achieve solids recoveries in excess of 80 percent. Highest solids capture using this polymer dose was 96 percent. Solids capture of 47-58 percent resulted from operation without polymer with feed concentrations of 34 g/ℓ and 65 g/ℓ. Feed concentration had no effect on solids recovery.

Highest attainable cake solids were 40-44 percent at low solids captures. At solids captures exceeding 80 percent, cake dry solids were 22-28 percent.

Phase 2 (High Ferric; with Polymer)

Results of all centrifugal dewatering runs conducted on sludge generated during Phase 2 are shown in Table 44 and on Figures 88-89.

For feed concentrations of 62 and 68 g/ℓ, a polymer dose of 0.36-0.52 weight percent was required to achieve solids captures of 90 percent and greater; highest capture using this polymer dose was 99 percent. Solids captures of 40-50 percent resulted from operation without polymer.

Inspection of Figure 88 shows that for one of the polymer doses there is no curve drawn. Some data scatter is apparent for the polymer range of 0.23-0.33 weight percent, but the points essentially fall between a lower and higher polymer range.

TABLE 42

DATA SUMMARY OF CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE FROM PHASE 3

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
1	10	44.6	3.62	0	26.6	38.3
2	10	44.6	5.76	0	25.3	36.7
3	25	44.6	3.50	0	27.6	38.1
4	25	44.6	5.66	0	26.6	36.2
5	20	44.6	3.65	0	27.3	37.7
6	20	44.6	6.06	0	25.6	36.6
7	29	44.6	11.9	0	27.0	34.0
8	20	44.6	15.3	0	25.7	34.7
9	20	44.6	4.72	0.10	20.1	37.8
10	20	44.6	5.79	0.15	16.2	98.2
11	20	44.6	6.74	0.11	18.0	53.5
12	20	44.6	11.7	0.10	19.3	45.6
13	20	44.6	3.64	0.25	16.8	91.3
14	20	44.6	6.20	0.24	17.5	92.8
15	20	44.6	8.89	0.22	17.0	83.7
16	20	44.6	11.6	0.21	16.1	74.2
17	20	44.6	6.74	0.34	18.7	95.7
18	20	44.6	8.23	0.39	18.6	99.3
19	20	44.6	11.5	0.32	18.0	93.6
20	20	44.6	15.0	0.30	18.0	85.5
21	20	44.6	6.06	0.40	19.0	96.3
22	20	44.6	9.73	0.38	18.6	96.0

(continued)

*Dow AP-30

TABLE 42, (continued)

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
23	20	44.6	11.2	0.39	17.7	97.5
24	20	44.6	15.1	0.40	18.8	86.6
25	20	47.2	4.04	0	26.9	35.0
26	20	47.2	6.20	0	27.9	32.5
27	20	47.2	11.6	0	27.2	31.3
28	20	47.2	15.2	0	26.7	28.4
29	20	47.2	3.91	0.13	20.7	39.7
30	20	47.2	6.06	0.13	18.2	86.0
31	20	47.2	9.03	0.12	17.7	65.4
32	20	47.2	11.6	0.11	19.7	48.3
33	20	47.2	3.77	0.25	17.2	93.7
34	20	47.2	6.20	0.25	18.0	93.3
35	20	47.2	8.76	0.24	16.6	91.9
36	20	47.2	11.5	0.24	17.1	78.8
37	20	47.2	6.60	0.36	18.4	97.0
38	20	47.2	8.49	0.36	18.5	95.2
39	20	47.2	11.7	0.24	17.4	82.9
40	20	47.2	14.4	0.33	18.0	79.4
41	20	47.2	6.00	0.65	18.7	96.2
42	20	47.2	8.89	0.42	18.9	96.8
43	20	47.2	12.0	0.47	18.3	92.9
44	20	47.2	15.5	0.36	18.8	87.1

*Dow AP-30

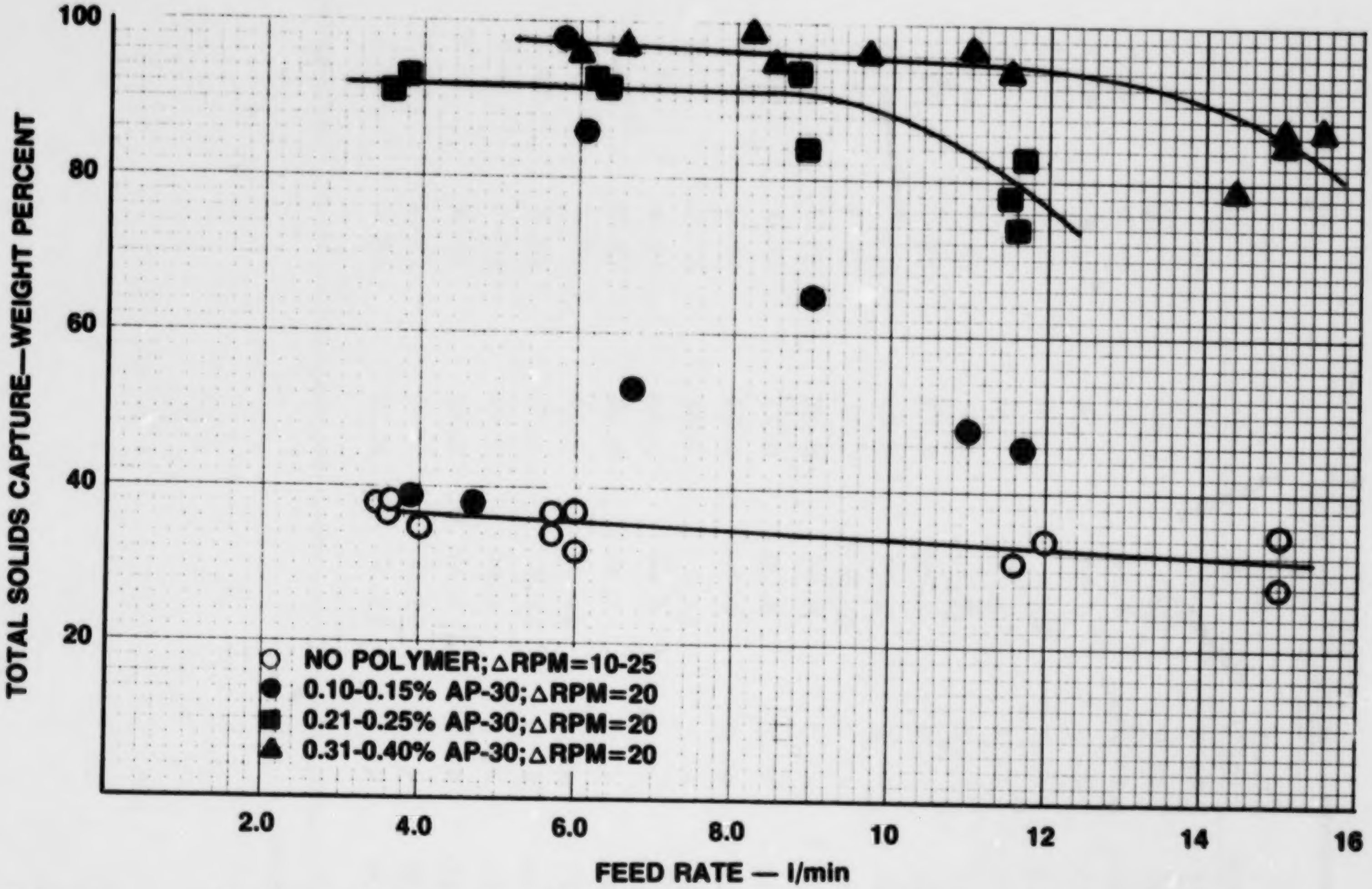

FIGURE 84. SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE — PHASE 3.
(Low Alum; with Polymer)

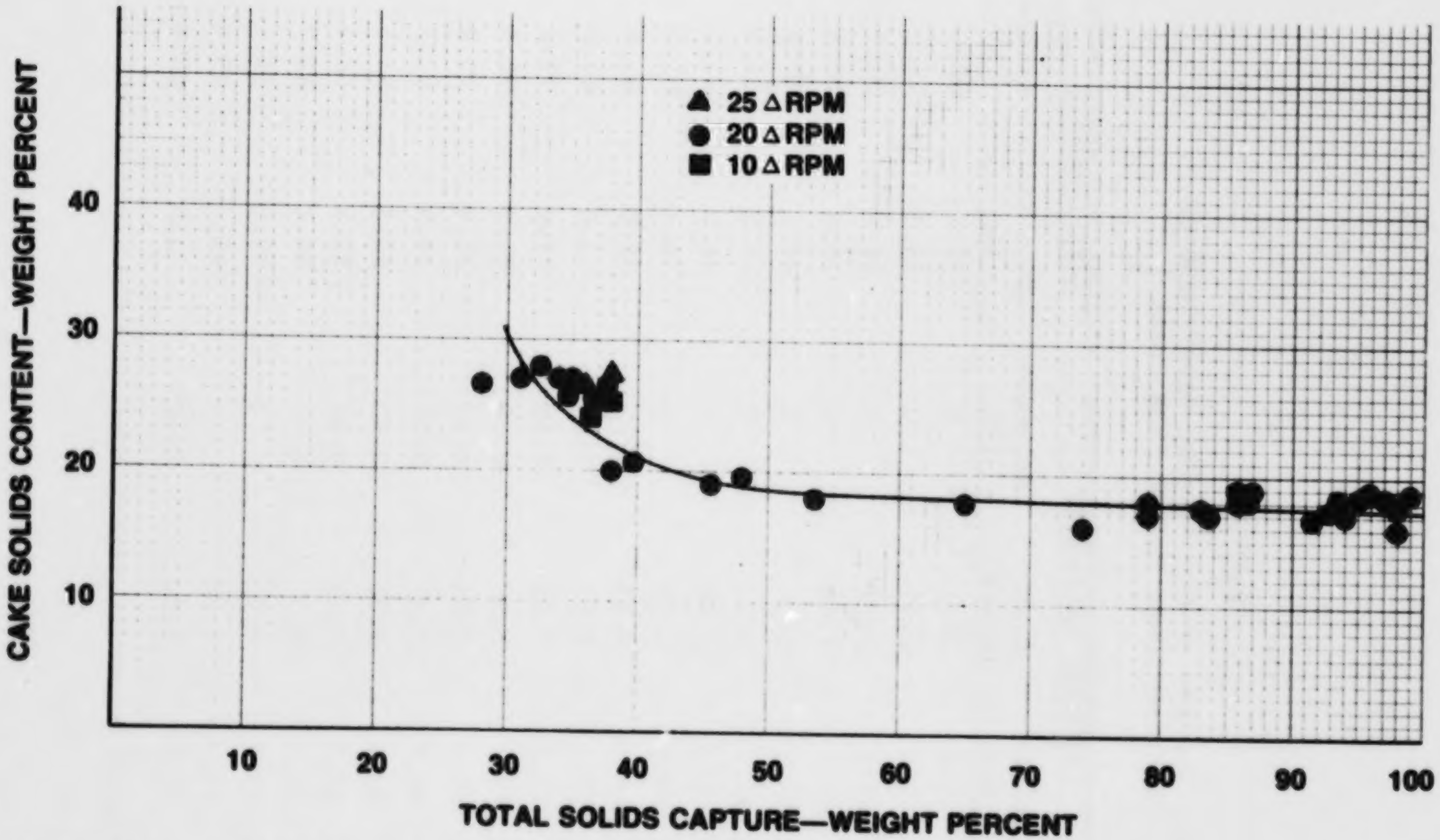

FIGURE 85. CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE — PHASE 3
(Low Alum; with Polymer)

TABLE 43

DATA SUMMARY OF CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE FROM PHASE 1

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
1	12	65.1	39.1	0	33.1	52.0
2	25	65.1	4.31	0	33.7	52.5
3	20	65.1	4.31	0	34.6	52.7
4	20	65.1	8.49	0	40.4	48.5
5	20	65.1	15.0	0	44.1	47.4
6	5	33.7	3.23	0	36.9	55.5
7	12	33.7	4.04	0	36.2	54.1
8	20	33.7	7.55	0	39.4	51.5
9	12	33.7	3.23	0.21	29.0	63.4
10	12	33.7	3.57	0.23	23.0	65.1
11	12	33.7	7.41	0.10	24.4	73.1
12	12	33.7	3.37	0.29	21.5	96.4
13	12	33.7	7.55	0.27	27.4	90.0
14	12	33.7	13.1	0.25	24.5	67.7
15	12	33.7	3.77	0.63	25.5	91.6
16	12	33.7	7.28	0.48	22.4	88.2
17	12	33.7	13.3	0.10	26.6	63.7
18	12	33.7	3.23	0	32.5	58.1
19	12	33.7	7.28	0	41.4	45.4
20	12	33.7	6.93	0.60	21.8	92.6

*Dow AP-30

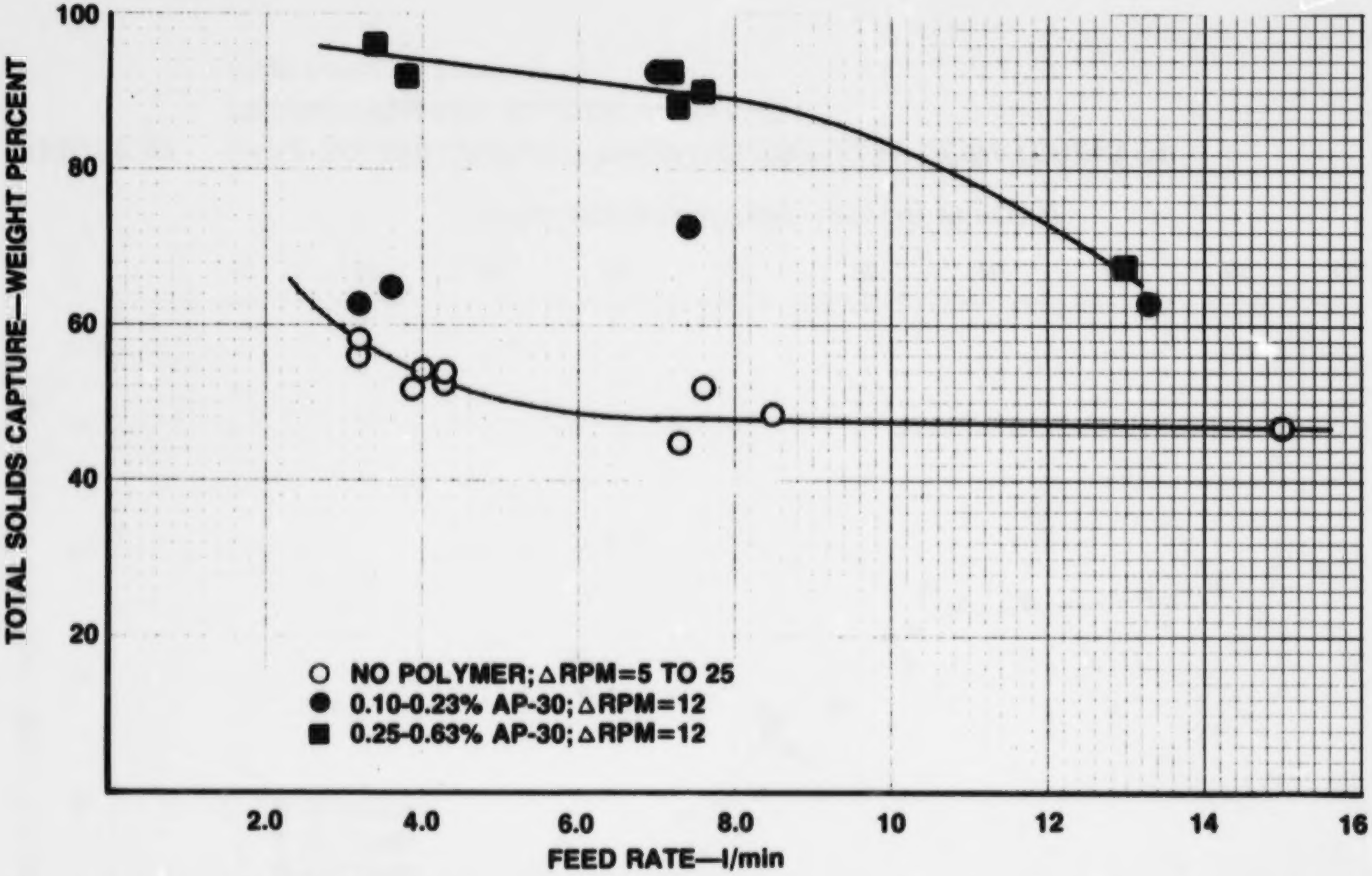

FIGURE 86. SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE — PHASE 1.
(High Ferric; no Polymer)

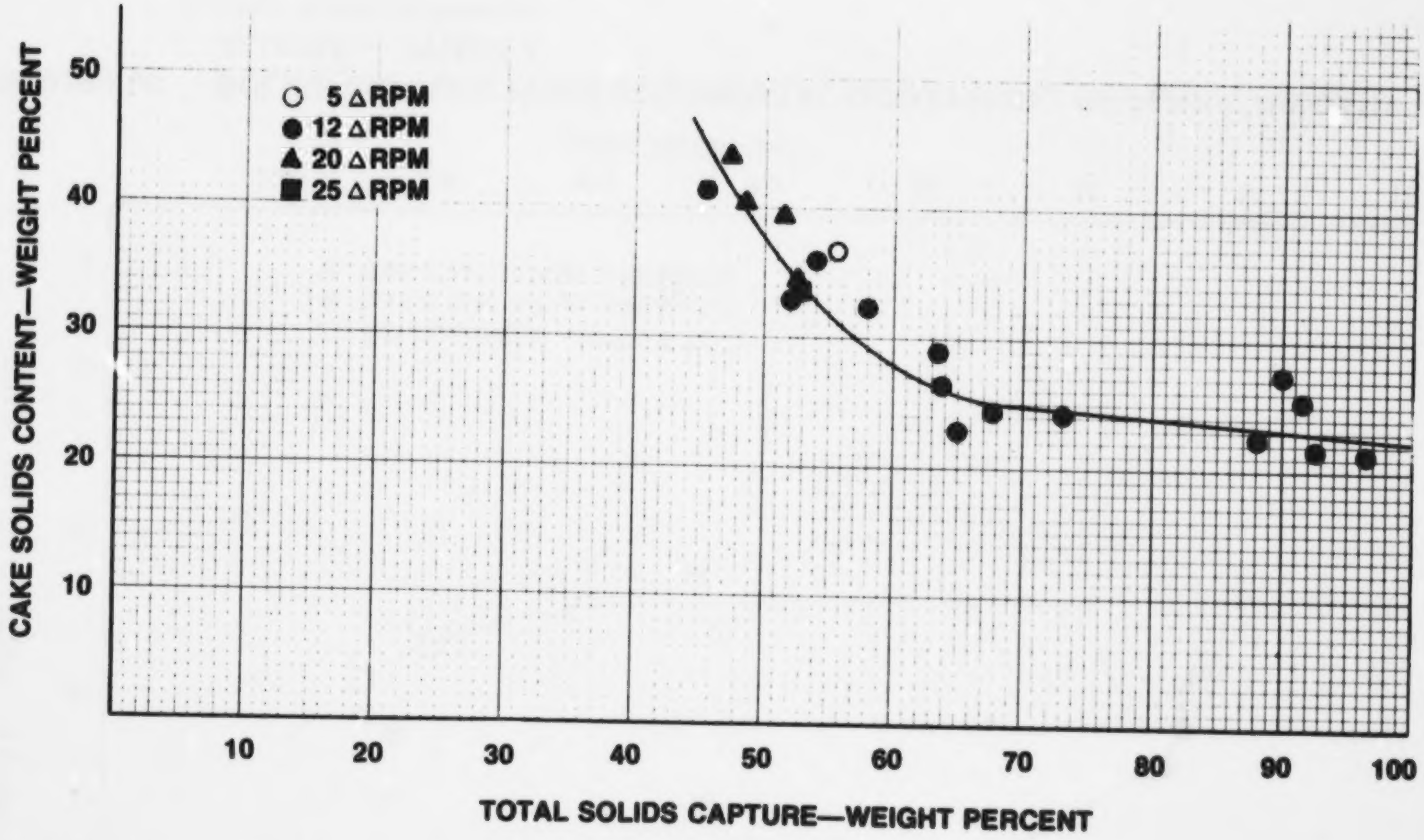

FIGURE 87. CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE — PHASE 1.
(High Ferric; no Polymer)

TABLE 44

DATA SUMMARY OF CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE FROM PHASE 2

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
1	10	61.6	3.77	0	31.7	53.3
2	10	61.6	4.04	0	32.7	52.4
3	25	61.6	3.64	0	34.5	49.2
4	25	61.6	6.60	0	32.4	49.2
5	20	61.6	2.91	0	33.4	53.3
6	20	61.6	5.93	0	32.1	50.9
7	20	61.6	12.0	0	33.4	47.4
8	20	61.6	14.2	0	33.3	43.8
9	20	61.6	3.59	0.11	31.4	49.5
10	20	61.5	6.20	0.13	28.5	58.8
11	20	61.6	8.49	0.11	27.6	50.9
12	20	61.6	11.9	0.11	29.3	54.8
13	20	61.6	3.77	0.26	25.8	77.4
14	20	61.6	6.06	0.23	24.6	71.0
15	20	61.6	8.66	0.18	27.5	62.0
16	20	61.6	6.87	0.19	27.6	48.4
17	20	61.6	5.79	0.33	25.4	47.1
18	20	61.6	8.35	0.30	25.6	81.0
19	20	61.6	11.6	0.27	26.8	73.8
20	20	61.6	14.8	0.25	27.8	52.1
21	20	61.6	6.40	0.43	26.7	90.9
22	20	60.7	8.09	0.43	25.5	86.9

*Dow AP-30

(continued)

TABLE 44. (continued)

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
23	20	60.7	11.5	0.39	26.4	72.7
24	20	60.7	15.3	0.35	27.3	61.0
25	10	67.8	3.37	0	32.0	48.8
26	10	67.8	6.06	0	31.9	49.1
27	25	67.8	3.91	0	31.0	48.0
28	25	67.8	6.74	0	31.4	44.7
29	20	67.8	3.91	0	32.5	49.3
30	20	67.8	6.20	0	30.3	46.8
31	20	67.8	11.7	0	30.7	42.5
32	20	67.8	15.4	0	31.9	40.6
33	20	67.8	3.91	0.11	26.4	53.0
34	20	67.8	6.20	0.13	25.9	42.3
35	20	67.8	9.30	0.12	26.7	43.0
36	20	67.8	11.6	0.14	27.2	41.6
37	20	67.8	3.64	0.24	23.0	88.3
38	20	67.8	6.33	0.29	23.0	88.8
39	20	67.8	8.22	0.25	23.2	80.5
40	20	67.8	11.7	0.23	24.0	61.9
41	20	67.8	5.93	0.41	24.2	92.0
42	20	67.8	8.76	0.36	25.0	91.4
43	20	67.8	12.5	0.33	25.7	89.8
44	20	67.8	15.2	0.28	24.9	76.1
45	20	67.8	6.06	0.52	25.6	93.0
46	20	67.8	9.26	0.45	24.8	96.8
47	20	67.8	7.28	0.36	23.5	98.8
48	25	67.8	3.64	0	31.8	46.5

*Dow AP-30

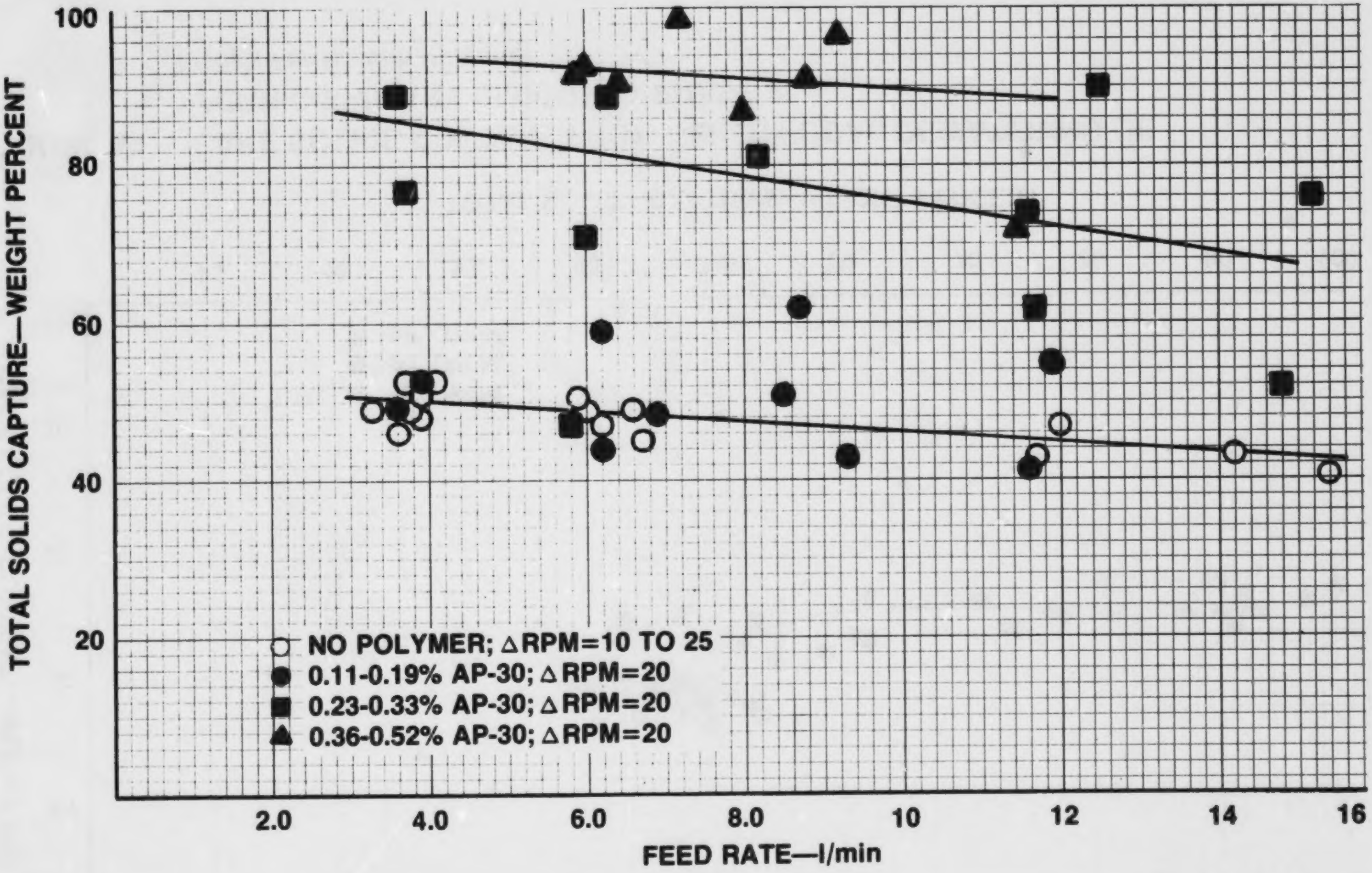

FIGURE 88. SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE — PHASE 2.
(High Ferric; with Polymer)

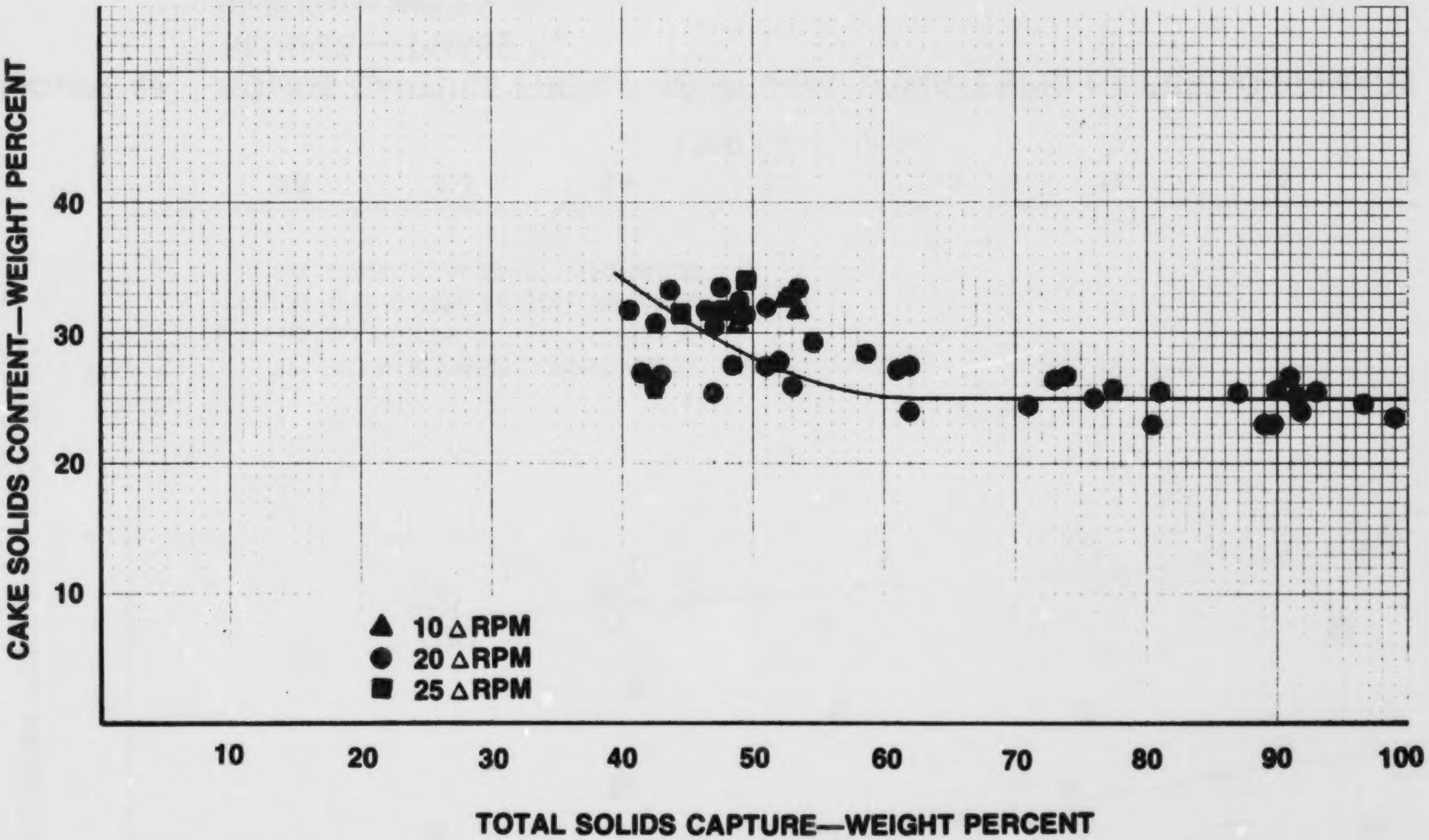

FIGURE 89. CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE — PHASE 2.
(High Ferric; with Polymer)

Highest cake solids attained was 35 percent at total solids captures of 40-50 percent. At captures in excess of 60 percent, cake solids were 24-26 percent.

Phase 3 (Low Ferric; with Polymer)

Results of all centrifuge runs conducted on ferric-primary sludge generated during Phase 3 are shown in Table 45 and Figures 90-91.

For a solids capture of 80 percent, a polymer dosage range of 0.20-0.30 weight percent was found to be necessary, while a polymer dosage of 0.34-0.58 weight percent resulted in solids captures greater than 90 percent. Highest solids capture was 94 percent at the polymer dose of 0.34-0.58 percent.

Solids captures of 45-50 percent was achieved without using polymer. Cake solids of 33-37 percent at low solids capture were generally achieved, while the cake solids decreased to 21-23 percent at captures greater than 80 percent.

PRIMARY SLUDGE

Results of all centrifugal dewatering runs conducted on primary sludge are shown in Table 46 and on Figures 92-93. From Figure 92, it is apparent that the use of anionic polymer was not effective in assisting centrifuge performance. This tends to indicate that a cationic polymer would have been the preferred conditioner. However, time prevented additional evaluations with cationic polymers.

Figure 92 shows a trend that was not present in any of the three prior phases (on either type of sludge). Regardless of the feed concentration (in this case 25 g/ℓ and 45 g/ℓ), and using no polymer, as the hydraulic loading decreased, solids captures in excess of 60 percent were achieved. Cake solids ranged from 25 percent at 44 percent solids capture to 20 percent at 90 percent solids capture.

The data generated during Phase 4 from primary sludge was included in an analysis of centrifugal dewatering performance to provide a clearer picture of the effects of chemical solids in the sludge on centrifugal dewatering performance.

DISCUSSION

Alum-Primary Sludge

Figures 94-95 illustrate the effect of the fraction of chemical solids in the sludge on the dewaterability of the sludge. On Figure 94, percent solids recovery, without polymer addition, was plotted versus hydraulic loading. Noting that Phases 1 and 2 represent sludge generated at virtually the same chemical

TABLE 45

DATA SUMMARY OF CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE FROM PHASE 3

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
1	10	46.6	3.91	0	37.1	50.9
2	10	46.6	6.20	0	35.9	48.5
3	25	46.6	3.77	0	33.4	49.4
4	25	46.6	5.93	0	33.0	49.5
5	20	46.6	3.77	0	34.6	49.5
6	20	46.6	6.20	0	32.8	46.9
7	20	46.6	11.7	0	34.0	46.6
8	20	46.6	15.2	0	35.5	44.2
9	20	46.6	3.77	0.17	21.5	78.1
10	20	46.6	6.33	0.14	22.2	74.5
11	20	46.6	8.62	0.13	23.0	73.5
12	20	46.6	11.5	0.13	24.1	64.5
13	20	46.6	3.78	0.25	22.0	93.8
14	20	46.6	6.21	0.28	20.9	76.7
15	20	46.6	8.51	0.27	22.5	70.5
16	20	46.6	11.3	0.20	21.9	76.6
17	20	46.6	6.16	0.42	22.6	88.0
18	20	46.6	9.18	0.38	22.7	75.9
19	20	46.6	11.8	0.37	22.5	75.2
20	20	46.6	15.3	0.36	23.5	70.3
21	20	46.6	6.23	0.58	22.6	89.0
22	20	46.6	8.37	0.55	21.2	86.6

(continued)

*Dow AP-30

TABLE 45. (continued)

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
23	20	46.6	11.6	0.49	22.6	82.7
24	20	46.6	13.6	0.34	25.0	73.9
25	20	58.4	3.38	0	34.6	49.1
26	20	58.4	6.21	0	35.7	46.5
27	20	58.4	11.8	0	35.8	46.0
28	20	58.4	15.1	0	36.9	44.3
29	20	58.4	3.85	0.04	24.8	65.2
30	20	58.4	6.21	0.11	23.3	84.5
31	20	58.4	8.78	0.12	23.5	70.0
32	20	58.4	11.9	0.15	26.4	68.8
33	20	58.4	3.59	0.23	20.9	89.7
34	20	58.4	6.08	0.28	21.3	88.8
35	20	58.4	8.37	0.27	24.6	82.1
36	20	58,4	11.9	0.22	23.6	77.6
37	20	58.4	8.24	0.37	22.3	87.7
38	20	58.4	8.64	0.37	22.5	79.8
39	20	58,4	9.86	0.29	24.0	72.2
40	20	58.4	15.0	0.35	24.9	74.6
41	20	58.4	6.21	0.54	20.5	93.0
42	20	58.4	8.64	0.48	22.1	90.0
43	20	58.4	12.3	0.37	23.2	83.1
44	20	58.4	15.3	0.30	23.2	78.6

*Dow AP-30

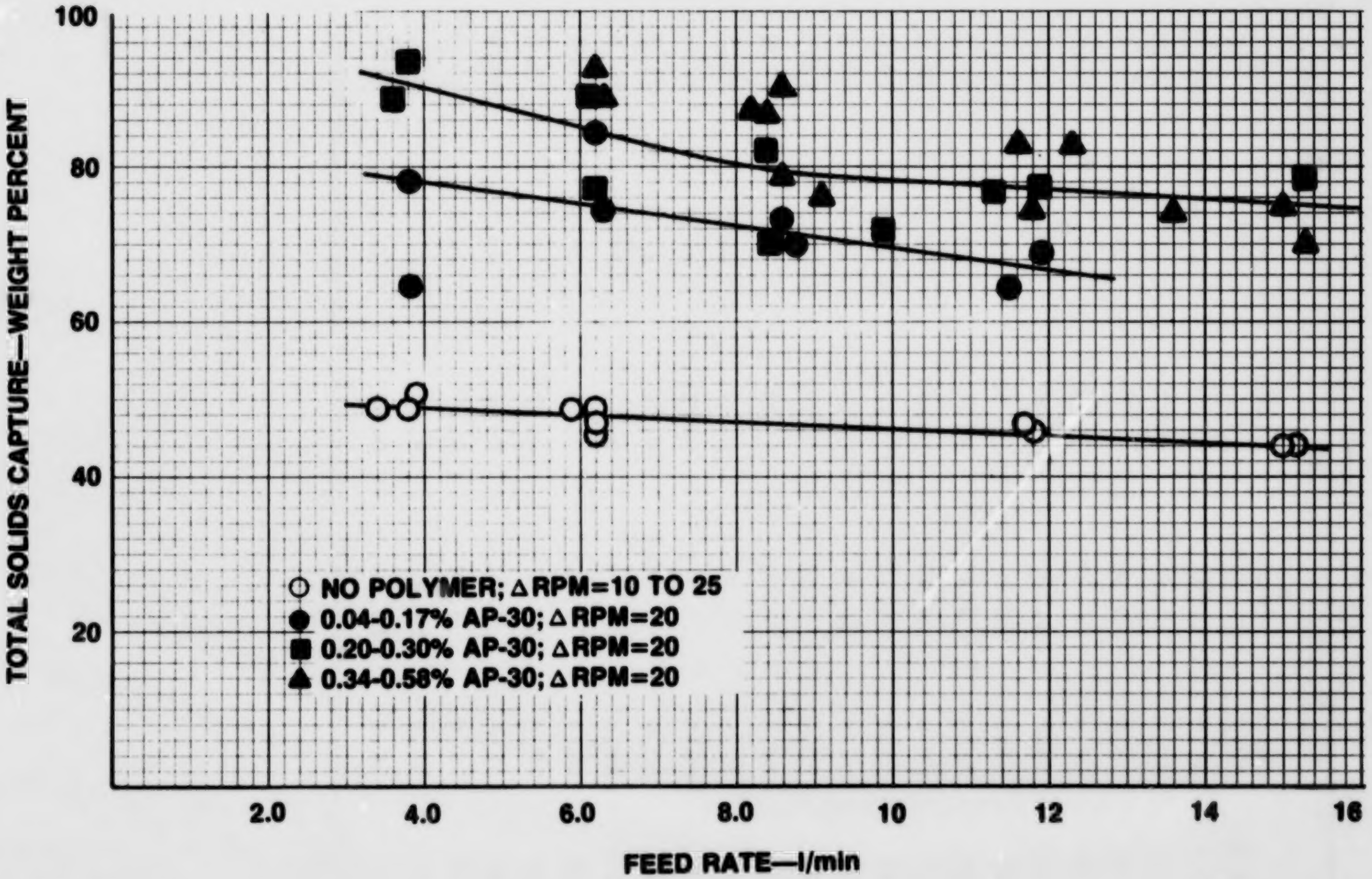

FIGURE 90.— SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE — PHASE 3.
(Low Ferric; with Polymer)

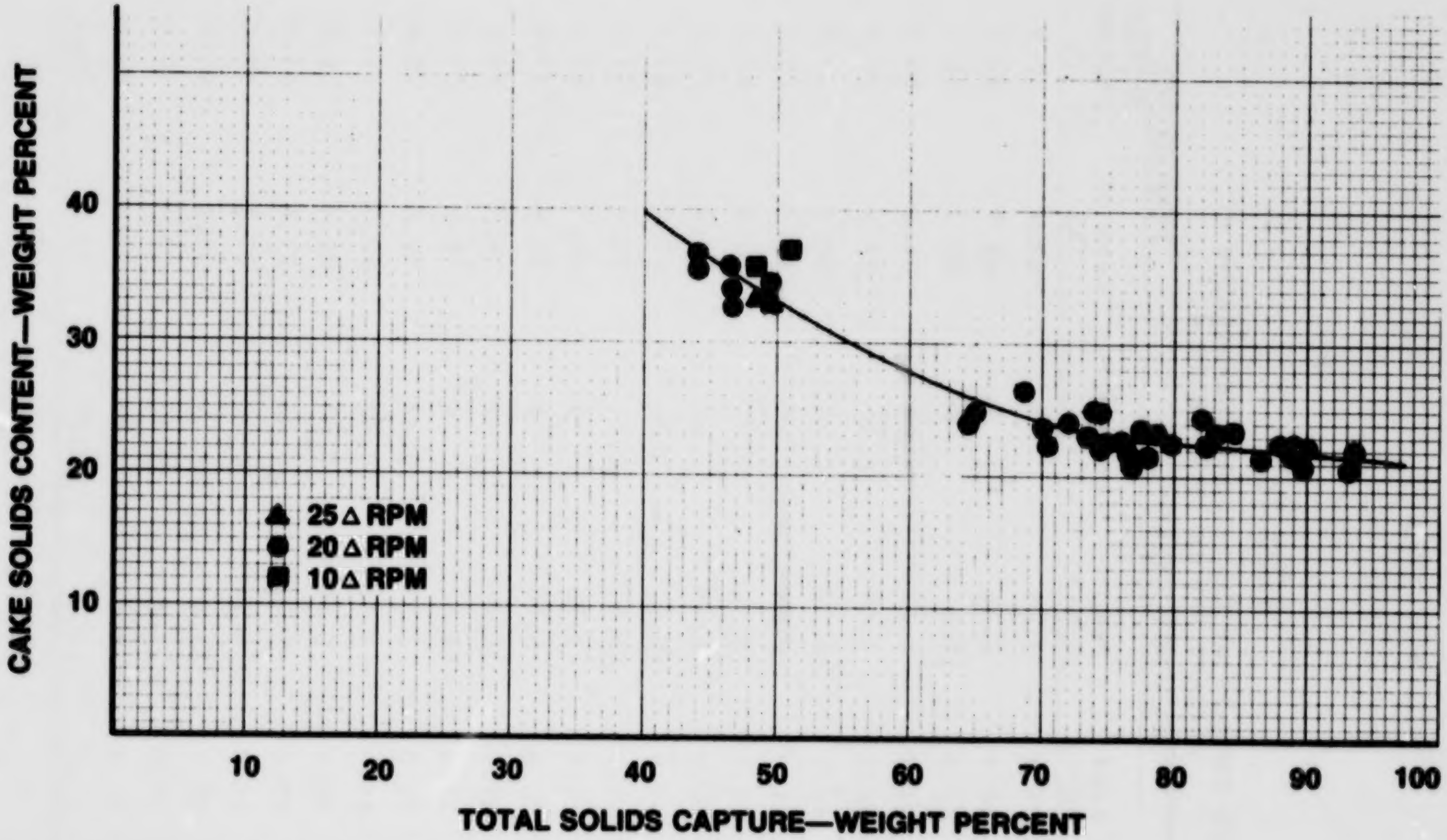

FIGURE 91. CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE — PHASE 3.
(Low Ferric; with Polymer)

TABLE 46

DATA SUMMARY OF CENTRIFUGAL DEWATERING OF PRIMARY SLUDGE FROM PHASE 4

		PROCESS VARIABLES			PERFORMANCE	
Run No.	ΔRPM	Feed (g/ℓ)	Feed (ℓ/min)	Polymer* (%)	Cake DS (%)	Solids Capture (%)
1	20	44.9	3.86	0	27.1	70.5
2	20	44.9	6.36	0	25.7	63.6
3	20	44.9	11.5	0	25.7	56.4
4	20	44.9	6.21	0.07	21.7	95.5
5	20	44.9	4.05	0.20	22.9	76.9
6	20	44.9	4.50	0.52	19.3	87.4
7	20	44.9	7.64	0.06	19.7	65.1
8	20	44.9	6.96	0.27	19.6	69.9
9	20	44.9	6.96	0.64	19.1	72.9
10	20	44.9	10.3	0.08	20.6	62.5
11	20	44.9	13.4	0.17	21.2	59.7
12	20	25.0	3.75	0	26.7	63.0
13	20	25.0	6.06	0	25.4	43.0
14	20	25.0	11.4	0	24.5	44.0
15	20	25.0	14.8	0	24.2	50.0
16	20	25.0	3.37	0.23	21.9	75.0
17	20	25.0	3.52	0.40	22.3	62.0
18	20	25.0	6.24	0.20	21.4	58.0
19	20	25.0	5.05	0.36	21.6	65.0
20	20	25.0	6.24	0.78	21.1	62.0
21	20	25.0	11.6	0.14	21.8	66.0
22	20	25.0	11.6	0.33	20.3	52.0
23	20	27.8	14.5	0.24	20.4	54.0
24	20	27.8	15.5	0.50	21.6	53.0

*Dow AP-30

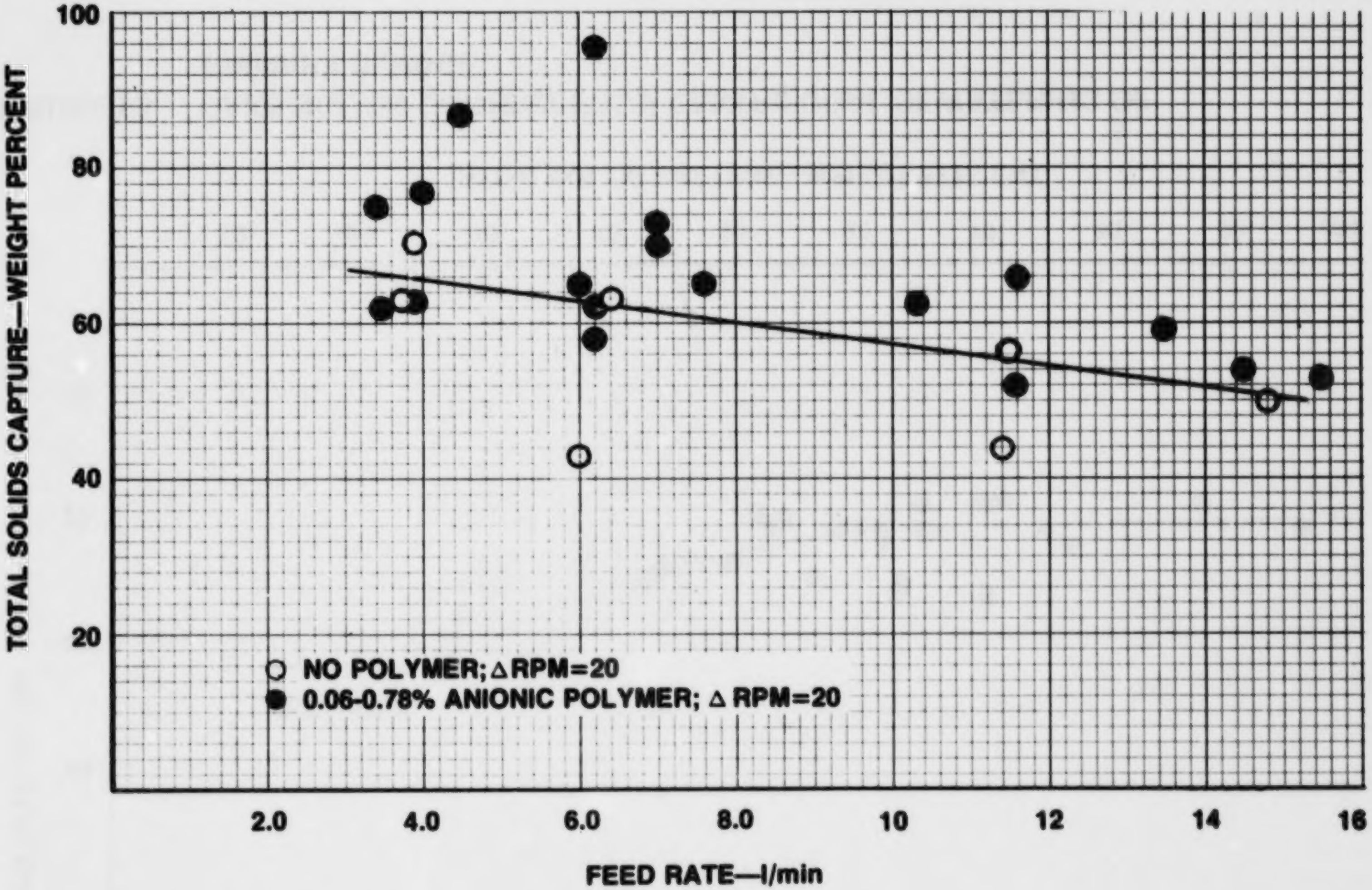

FIGURE 92. SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF PRIMARY SLUDGE

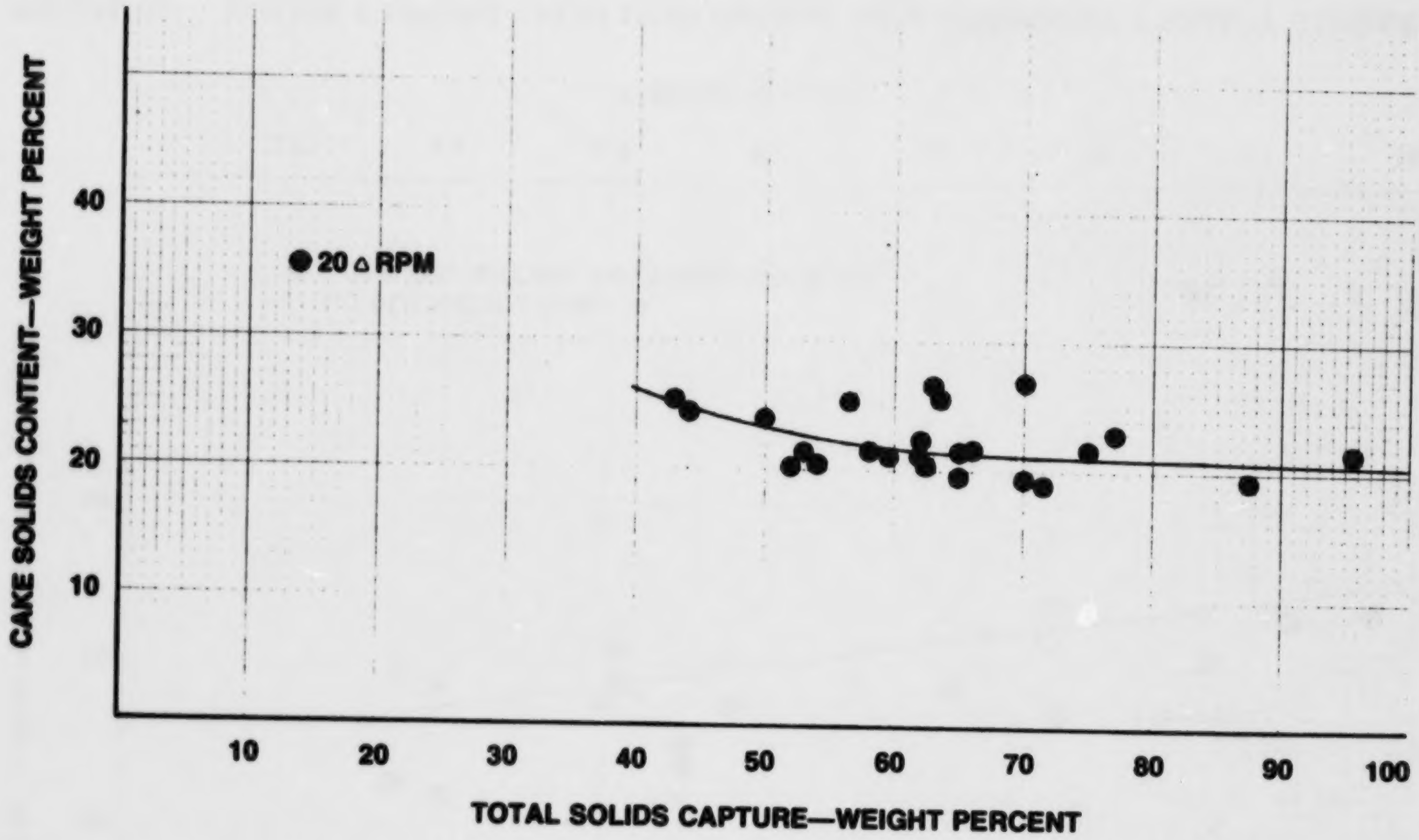

FIGURE 93. CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF PRIMARY SLUDGE.

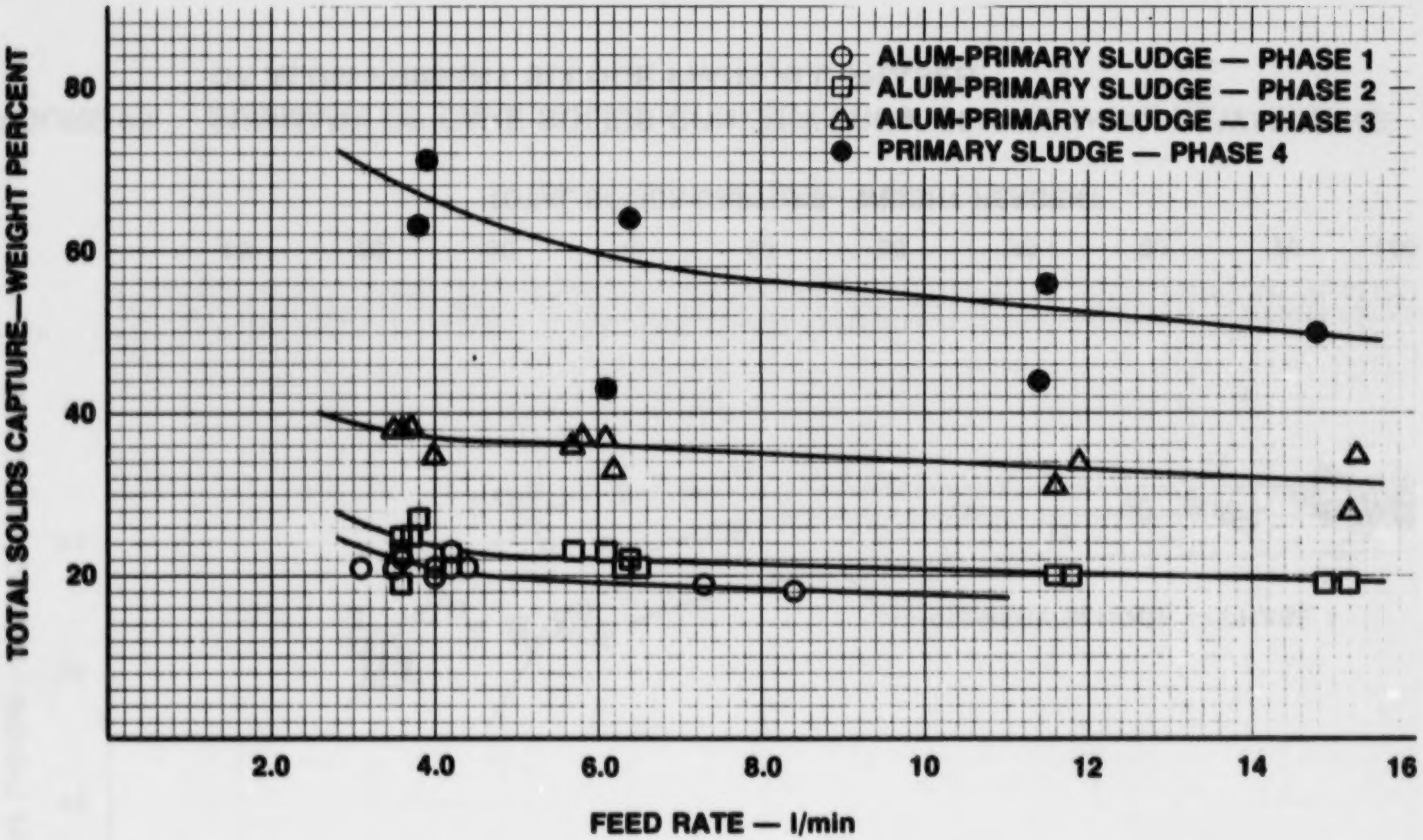

FIGURE 94. SUMMARY OF SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE FROM ALL PHASES.

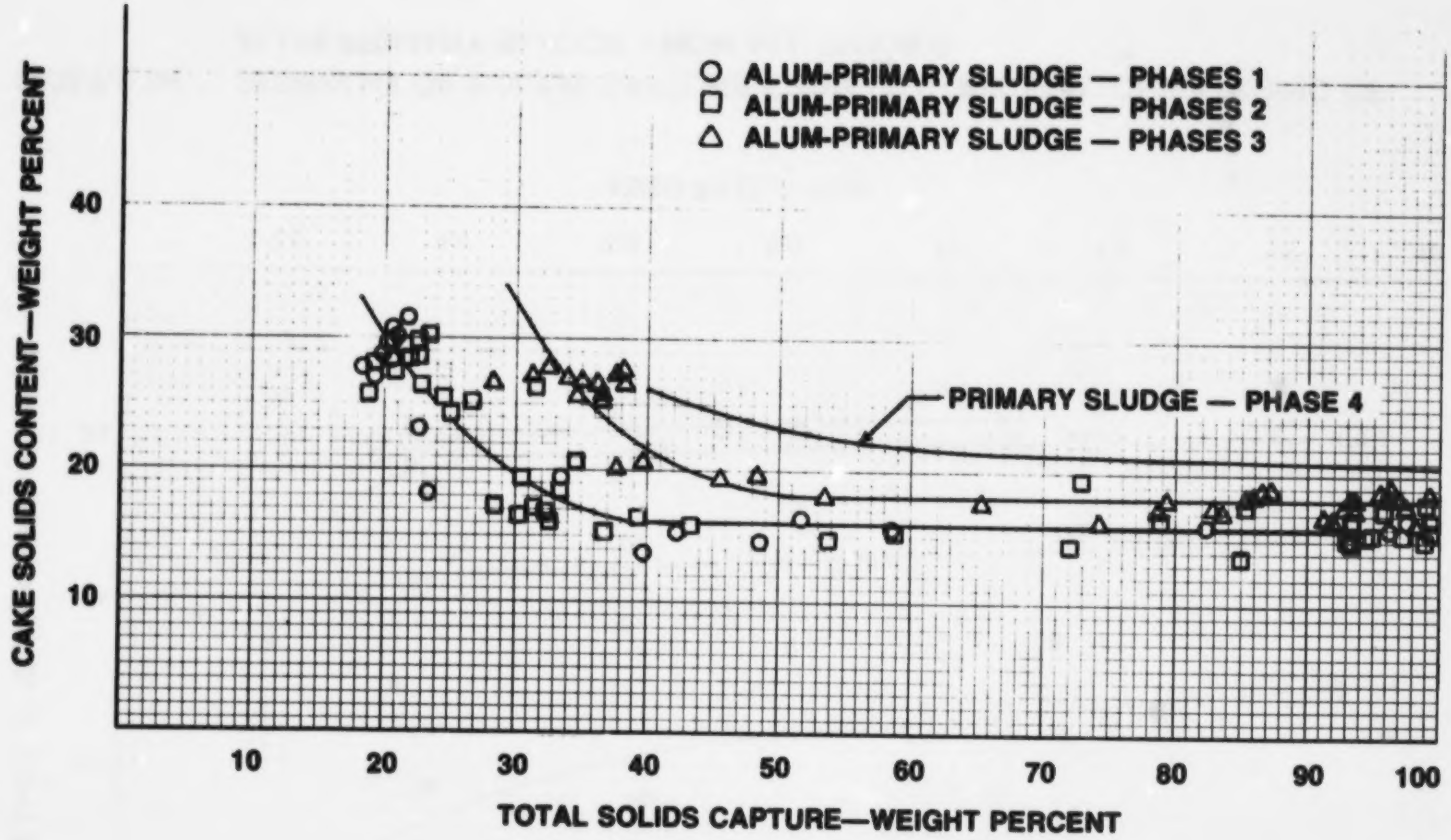

FIGURE 95. SUMMARY OF CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE FROM ALL PHASES.

dosage, that Phase 3 represents sludge generated at a lower chemical dosage, and that Phase 4 represents primary sludge, it was found that as the fraction of chemical solids present in the sludge decreased, the maximum percent solids recovery increased for any given hydraulic loading. Also, note that there was little difference in the solids recovery in the first two phases where the chemical solids fraction in the sludges were virtually the same. All results shown on Figure 94 represent tests performed using no polymer during centrifugation.

A similar relationship was observed for cake solids content, as shown on Figure 95. As the fraction of chemical solids present in the sludge was reduced, cake solids content increased for a given solids capture. Results from Phase 1 and Phase 2 were nearly identical.

Figure 96 shows how the maximum observed solids recovery drastically decreased as the fraction of chemical solids in the sludge increased. Again note the little difference in the results of Phases 1 and 2. All results shown on Figure 95 represent tests conducted on the centrifuge using no polymer.

Figure 97 is a graphical illustration of the effect of the chemical solids fraction on cake solids content. The summary of results plotted on Figure 97 shows that, for any level of solids capture, cake solids content decreased as the fraction of chemical solids in the alum-primary sludge increased. The effect of chemical solids fraction on cake solids content was especially apparent at low levels of solids capture. At high levels of solids capture, cake solids content was relatively insensitive to the fraction of chemical solids present in the sludge since all sludge constituents were efficiently captured and exhibited themselves in the cake. At lower levels of solids capture, however, cake solids content was sensitive to the fraction of chemical solids present in the feed sludge due to the classification characteristics of the solid bowl centrifuge.

Ferric-Primary Sludge

Figures 98-99 show the results of all centrifuge runs performed on ferric-primary sludge using no polymer. All curves for ferric-primary sludge shown on Figure 98 are flat with little difference between them. The results obtained for primary sludge were plotted on Figure 98 and indicated that solids recovery for primary sludge was generally higher than for ferric-primary sludge at equivalent centrifuge loadings.

Figure 99 shows cake solids content versus solids capture. Again, there is little difference in the results of all three ferric-primary sludges; at 40-45 percent solids captures, cake solids of 40 percent attained. In all cases, cake solids

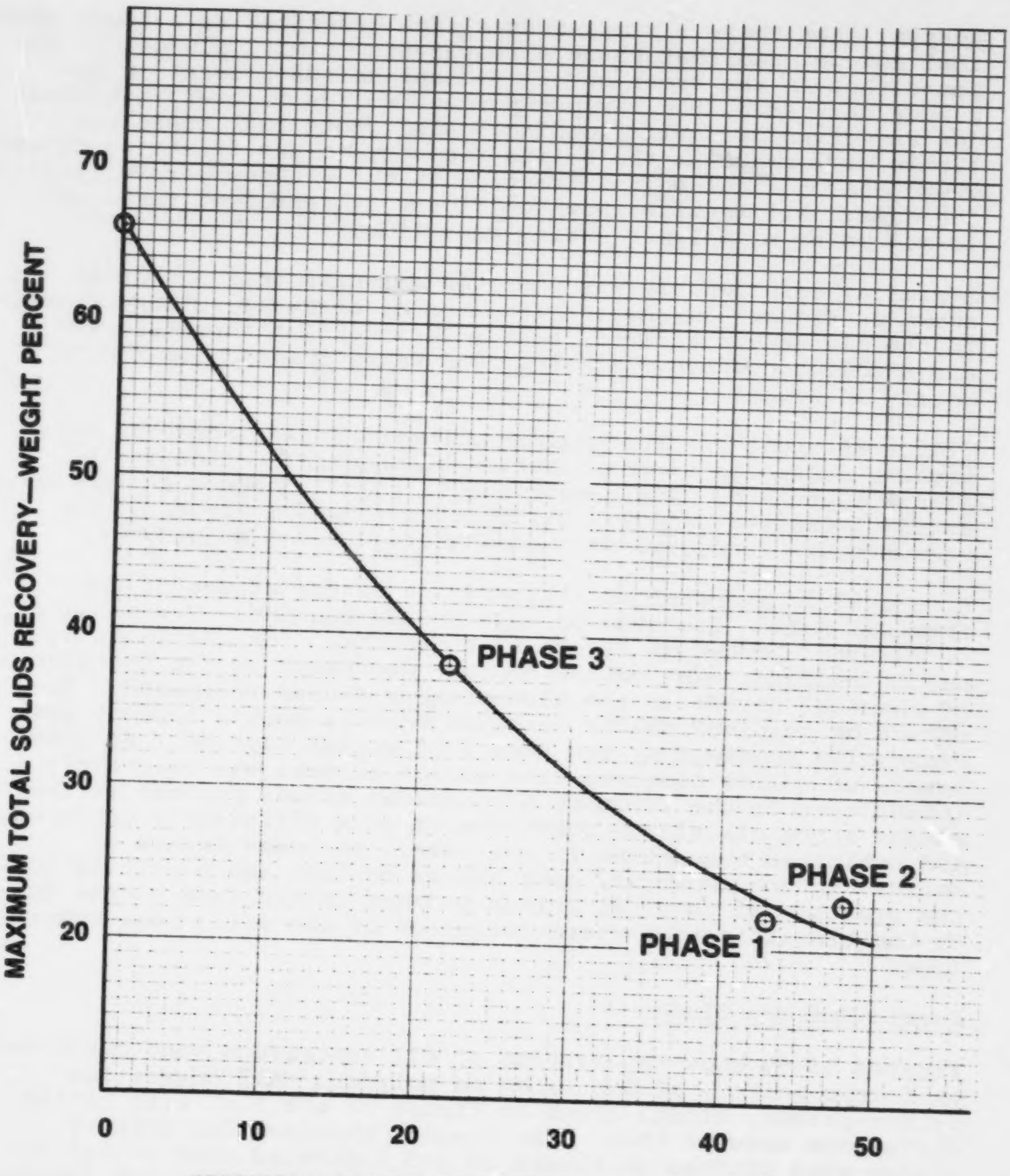

FIGURE 96. EFFECT OF CHEMICAL SOLIDS CONTENT OF ALUM-PRIMARY SLUDGE ON MAXIMUM TOTAL SOLIDS RECOVERY WITHOUT POLYMER.

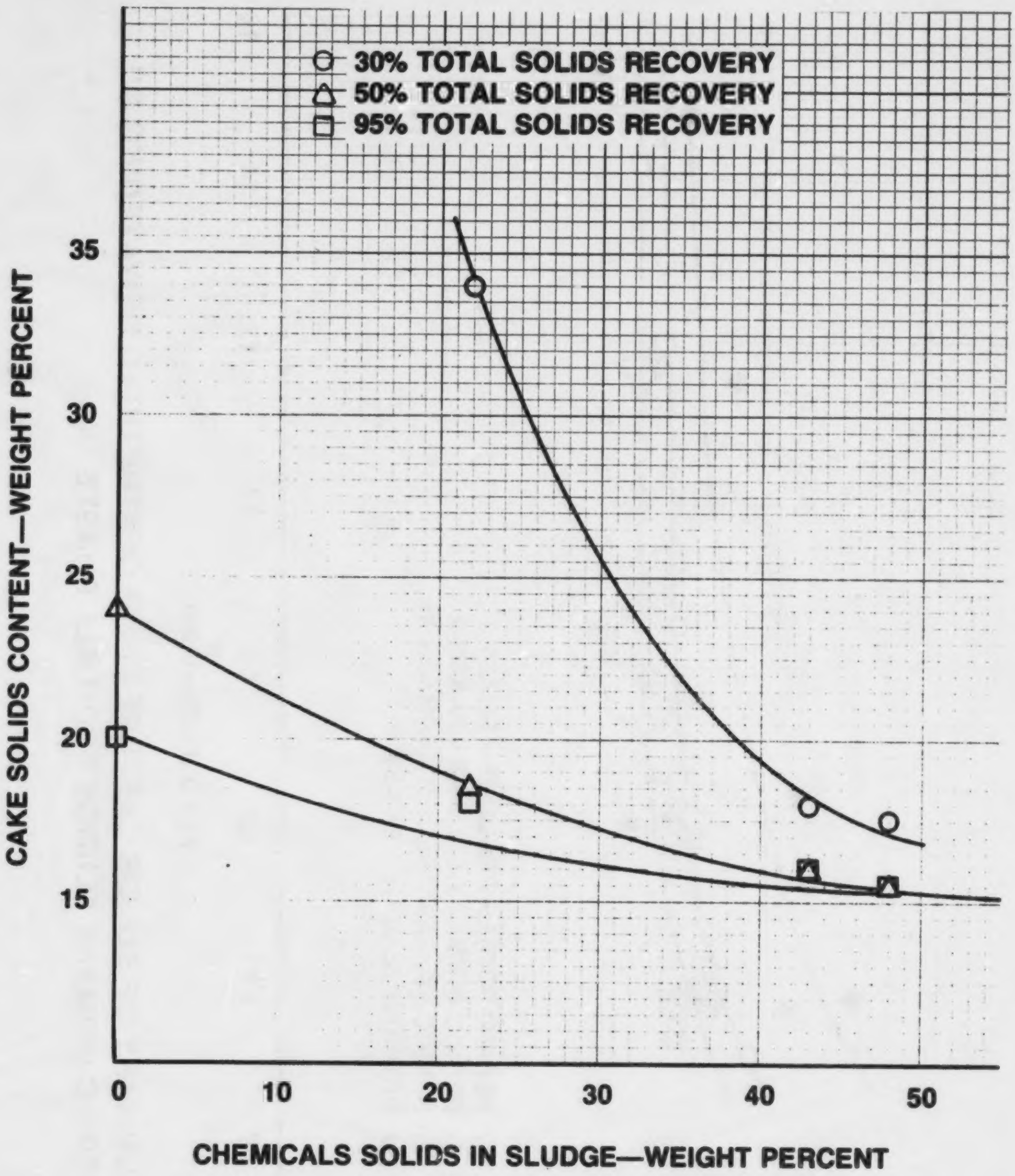

FIGURE 97. EFFECT OF CHEMICAL SOLIDS CONTENT OF ALUM-PRIMARY SLUDGE ON CAKE SOLIDS CONTENT.

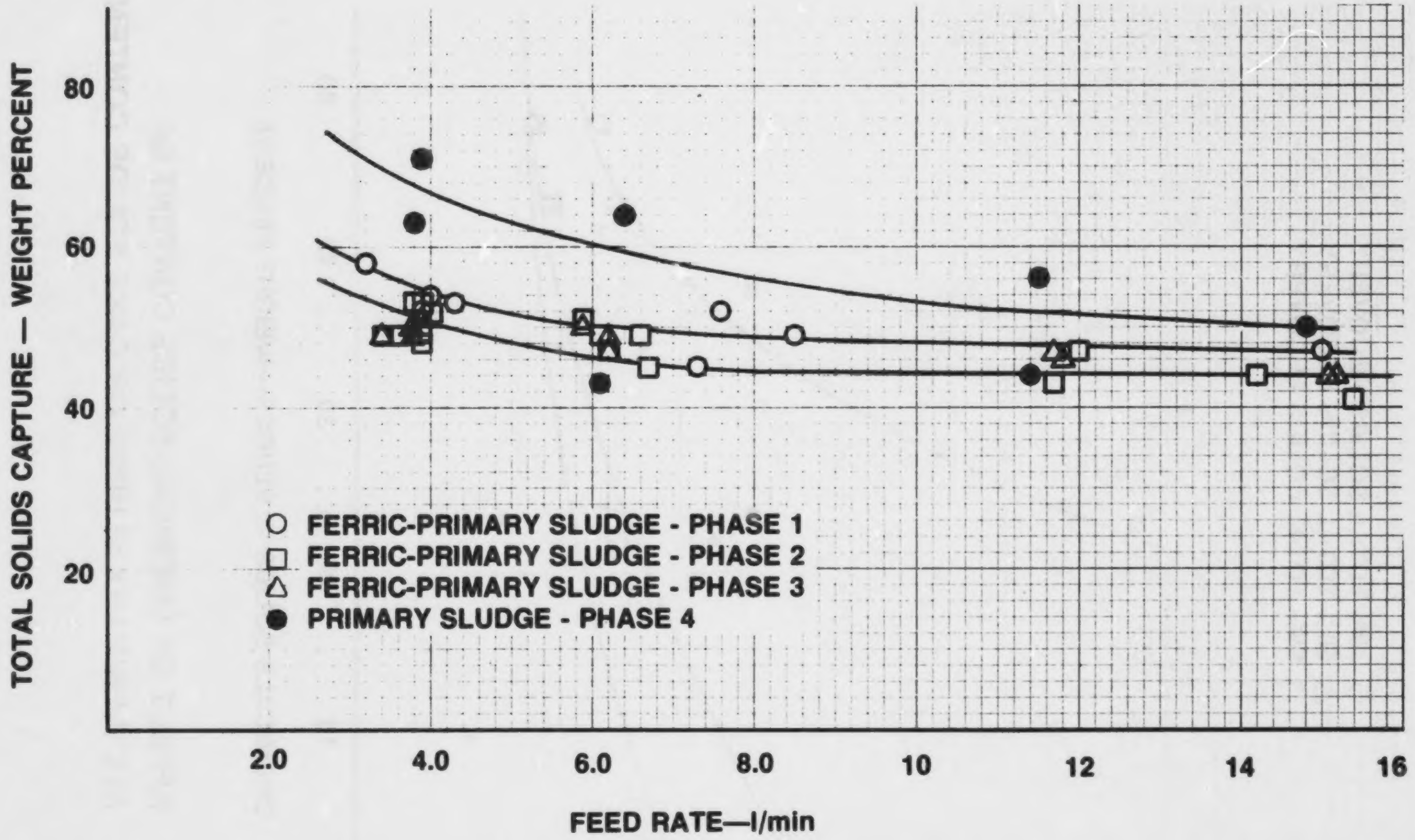

FIGURE 98. SUMMARY OF SOLIDS CAPTURE FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE FROM ALL PHASES.

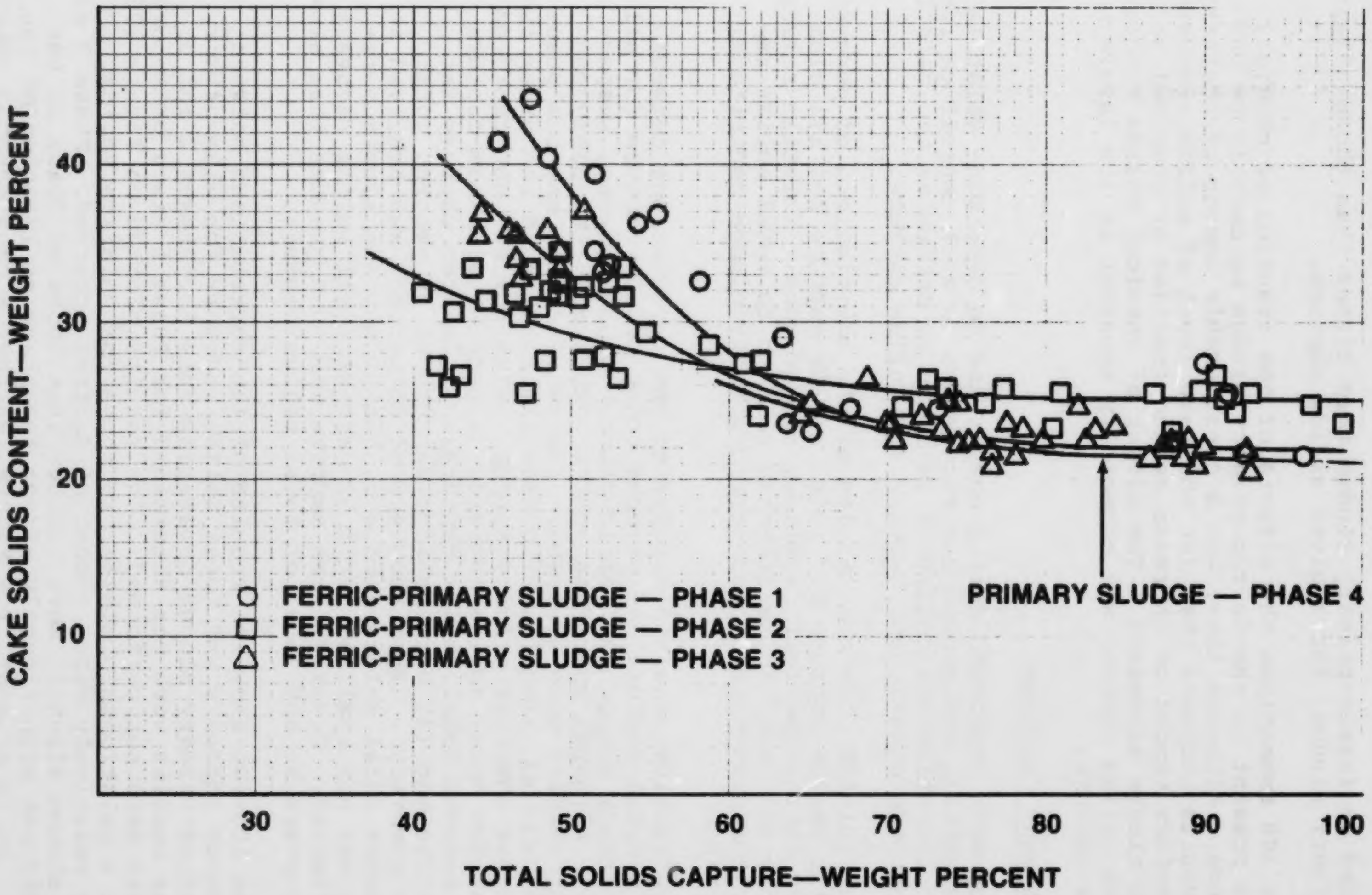

FIGURE 99. SUMMARY OF CAKE SOLIDS CONTENT FROM CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE FROM ALL PHASES.

attained on ferric-primary sludge were higher than those attained on primary sludge, for a given solids capture.

Figure 100 summarizes the effects of the fraction of chemical solids present in the ferric-primary sludge on centrifuge performance. Although there was a considerable amount of scatter, the results indicate that (at the same level of solids recovery) cake solids tended to increase as the fraction of chemical solids in the sludge increased. The effect of chemical solids fraction on cake solids content was especially apparent at low levels of solids capture.

Constituent Captures

Studies were conducted during each phase to determine volatile solids, metal, and phosphorus recoveries on each centrifuge run. This information was collected in order to determine the relationship between constituent capture and solids capture.

Figures 101-102 show the results of the correlations of percent volatile solids capture and total solids capture for all alum-primary sludge and ferric-primary sludge runs. Volatile solids recovery was found to be directly related to total solids recovery for alum-primary and ferric-primary sludge centrifugal dewatering.

Figures 103-104 show the results of the correlations between percent metal capture and percent total solids capture for all alum-primary sludge and ferric-primary sludge phases, respectively. Although some data scatter occurred, reasonably consistent results were achieved. The results from any one phase did not differ significantly from the results of the other two for either chemical-primary sludge. Aluminum was not captured well at low total solids captures (6 percent aluminum capture at 20 percent total solids capture), while iron capture was more efficient (35 percent iron capture at 20 percent total solids recovery). Iron recovery of 90 percent was achieved at 70 percent total solids recovery, but 90 percent aluminum recovery was not realized until 92 percent total solids recovery was effected. Clearly, iron capture was much greater than aluminum capture at any level of total solids capture.

Figures 105-106 show the results of the correlations between phosphorus recovery and total solids recovery for alum-primary and ferric-primary sludge dewatering runs for Phases 2 and 3. Similar results were not determined for Phase 1 since phosphorus analyses were not performed. The results plotted on Figure 105 showed a relationship similar to those for volatile solids. Only a relatively minor amount of scatter occurred. Phase 3 results showed slightly more scatter than those of Phase 2, but this did not significantly affect the shape or location of the curve. It is evident from Figure 106 that a considerable amount

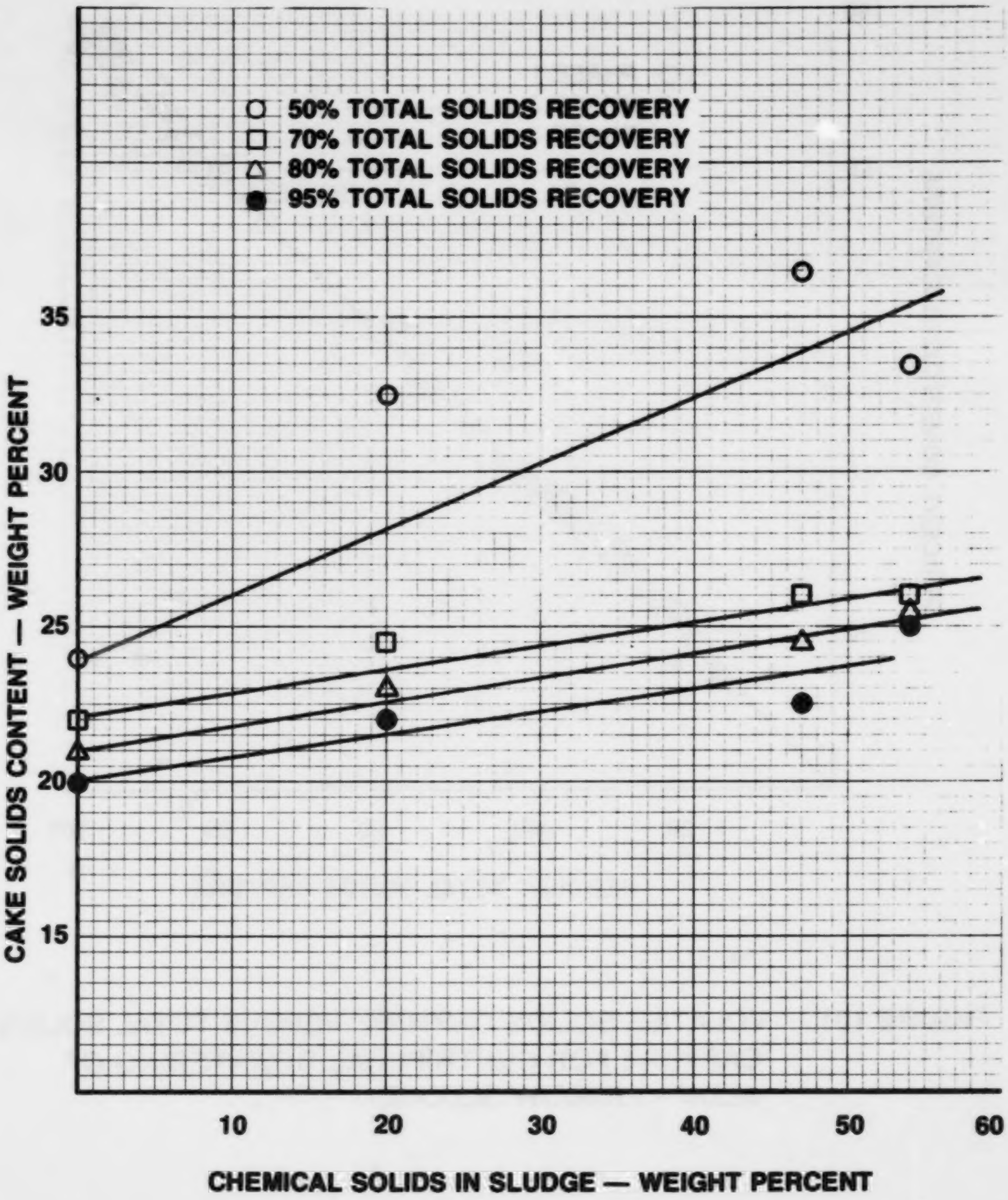

FIGURE 100. EFFECT OF CHEMICAL SOLIDS CONTENT OF FERRIC-PRIMARY SLUDGE ON CAKE SOLIDS CONTENT.

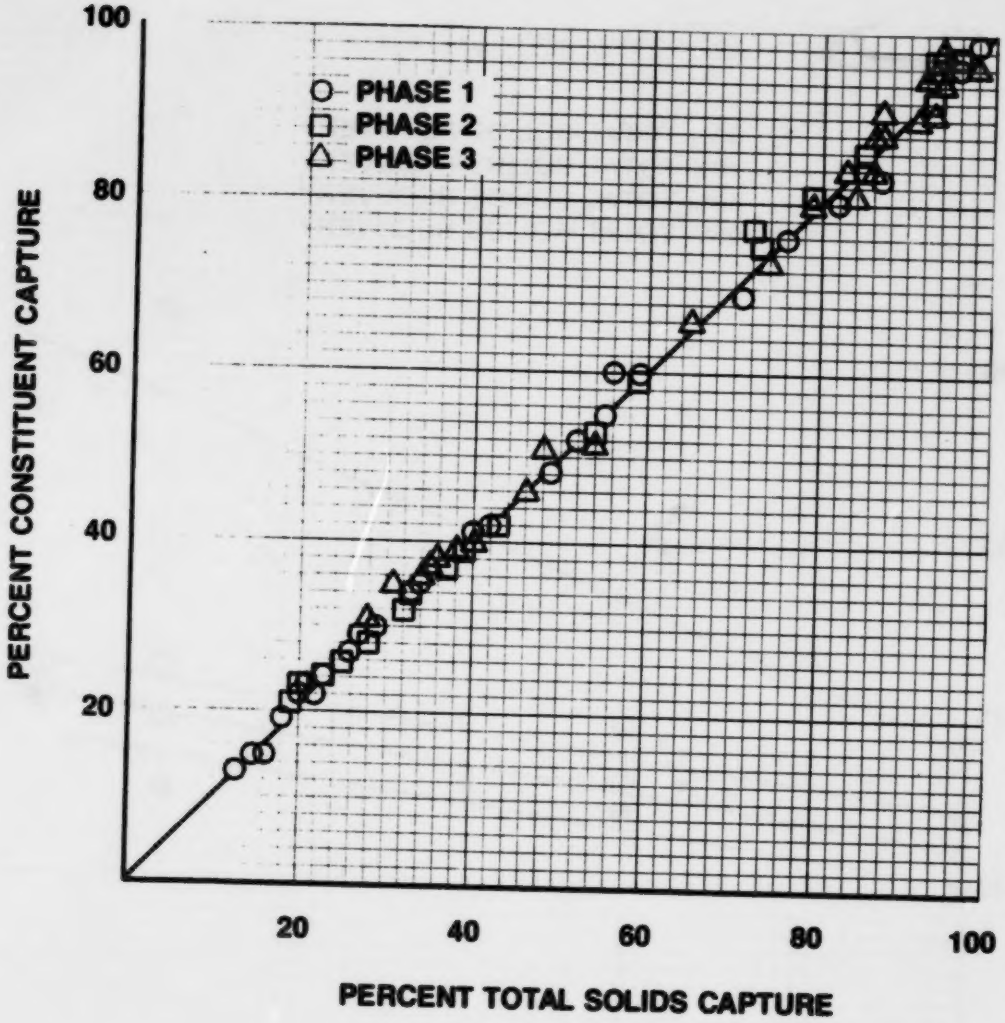

FIGURE 101. VOLATILE SOLIDS CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF ALUM - PRIMARY SLUDGE.

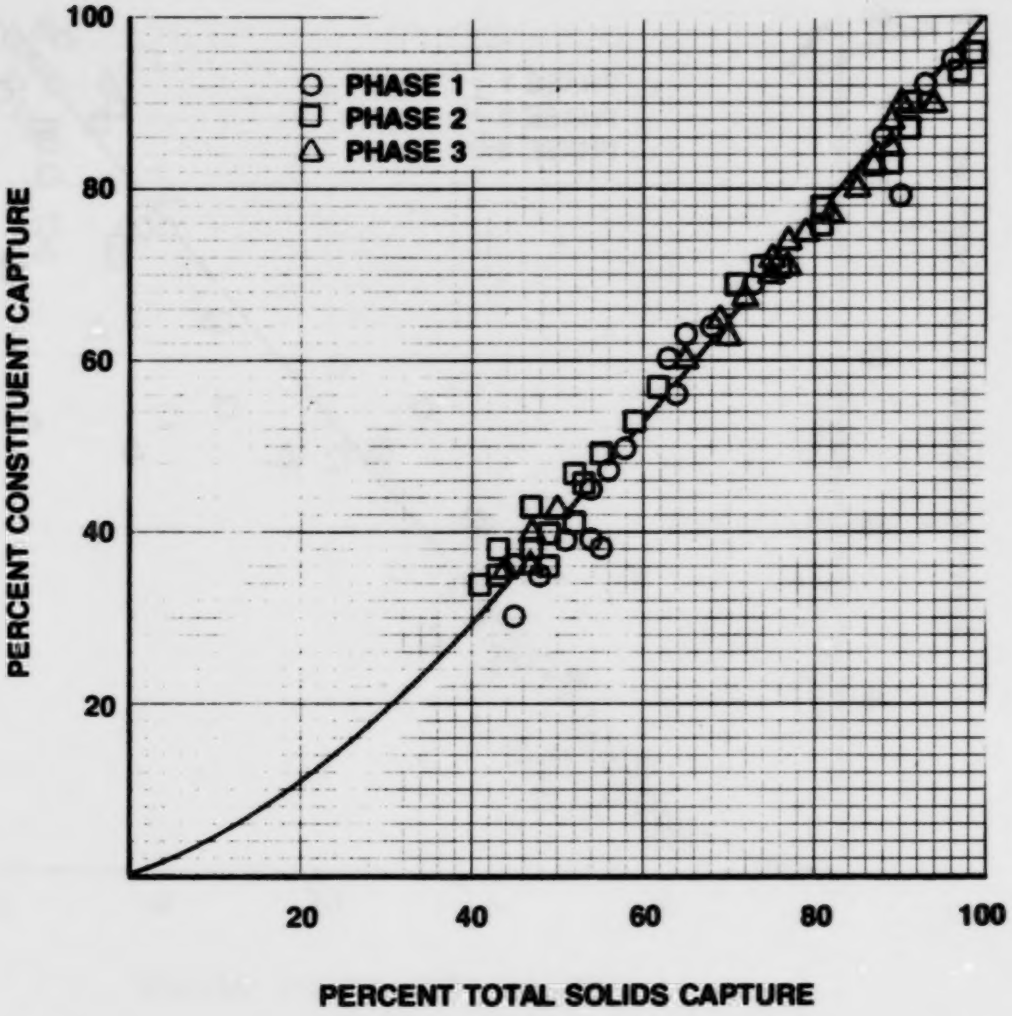

FIGURE 102. VOLATILE SOLIDS CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE.

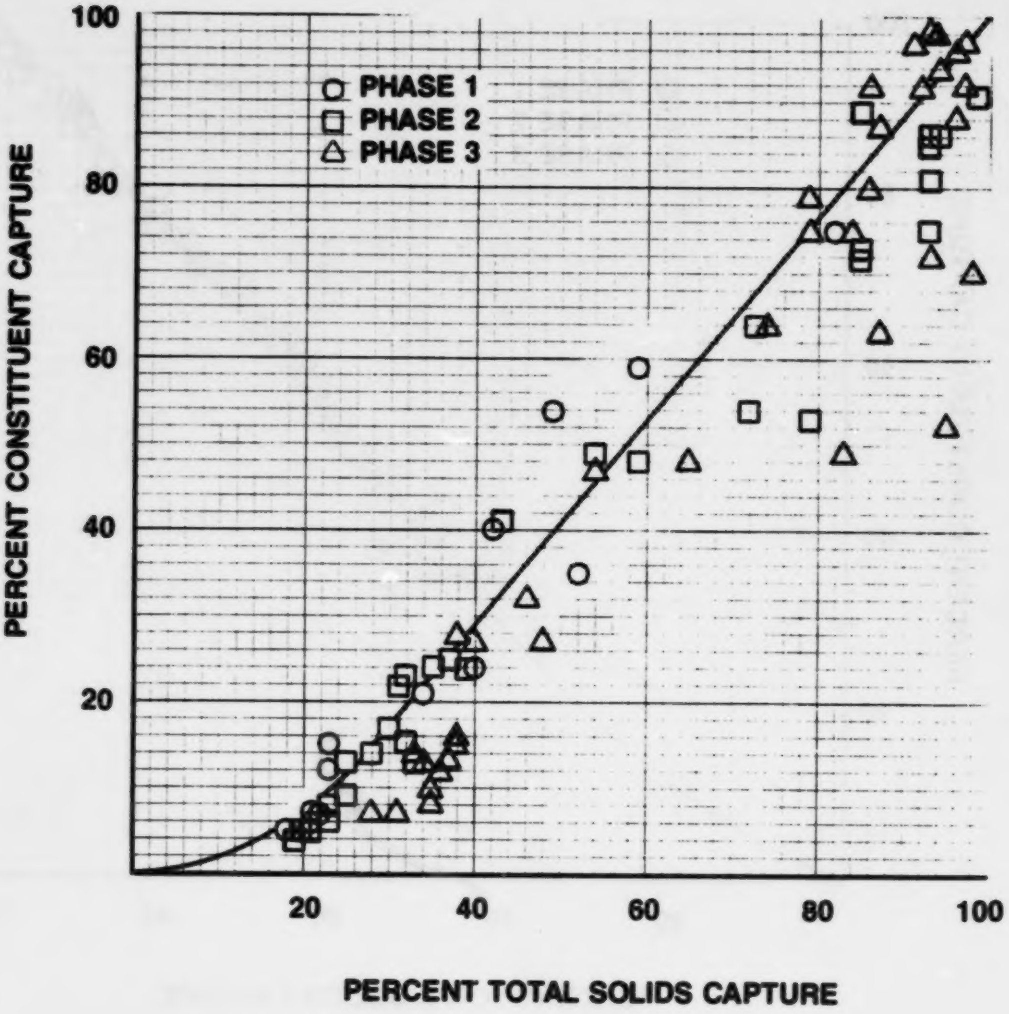

FIGURE 103. ALUMINUM CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE.

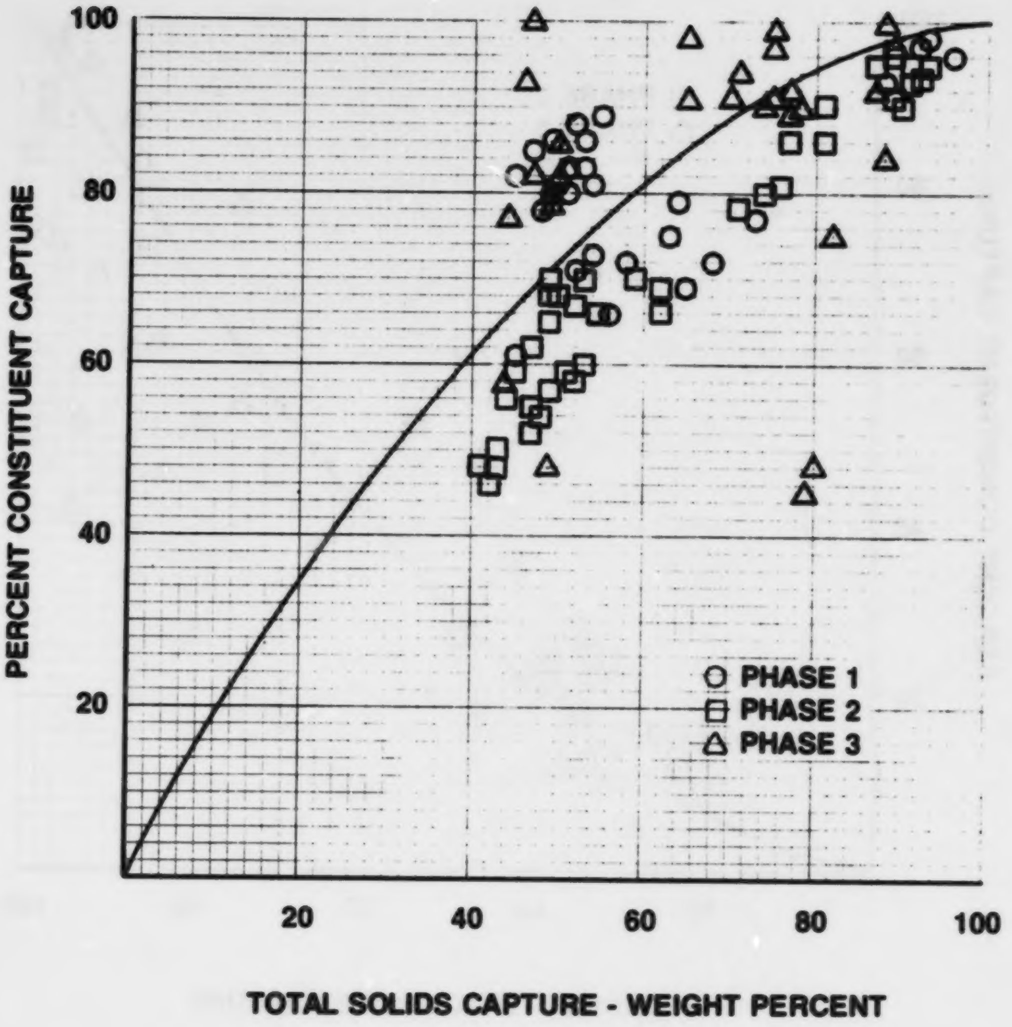

FIGURE 104. IRON CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE.

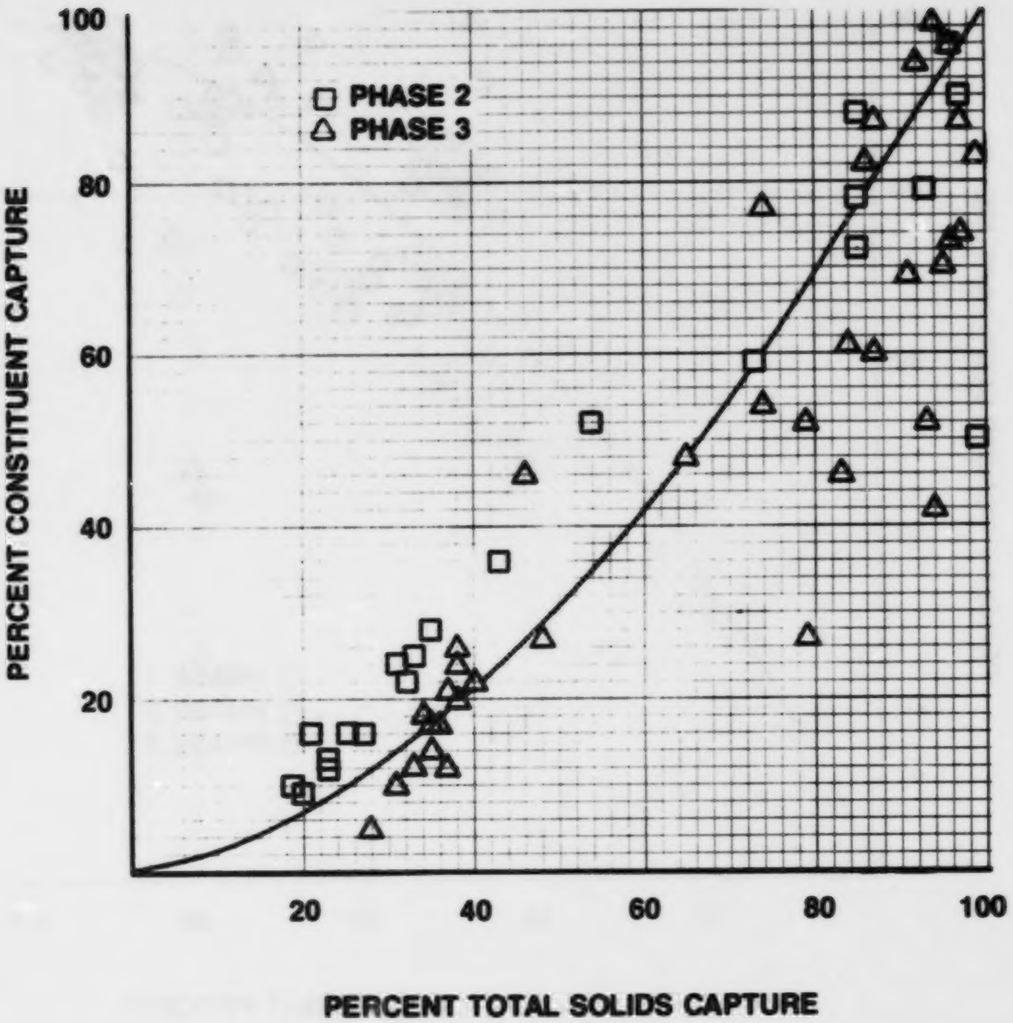

FIGURE 105. PHOSPHORUS CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF ALUM-PRIMARY SLUDGE.

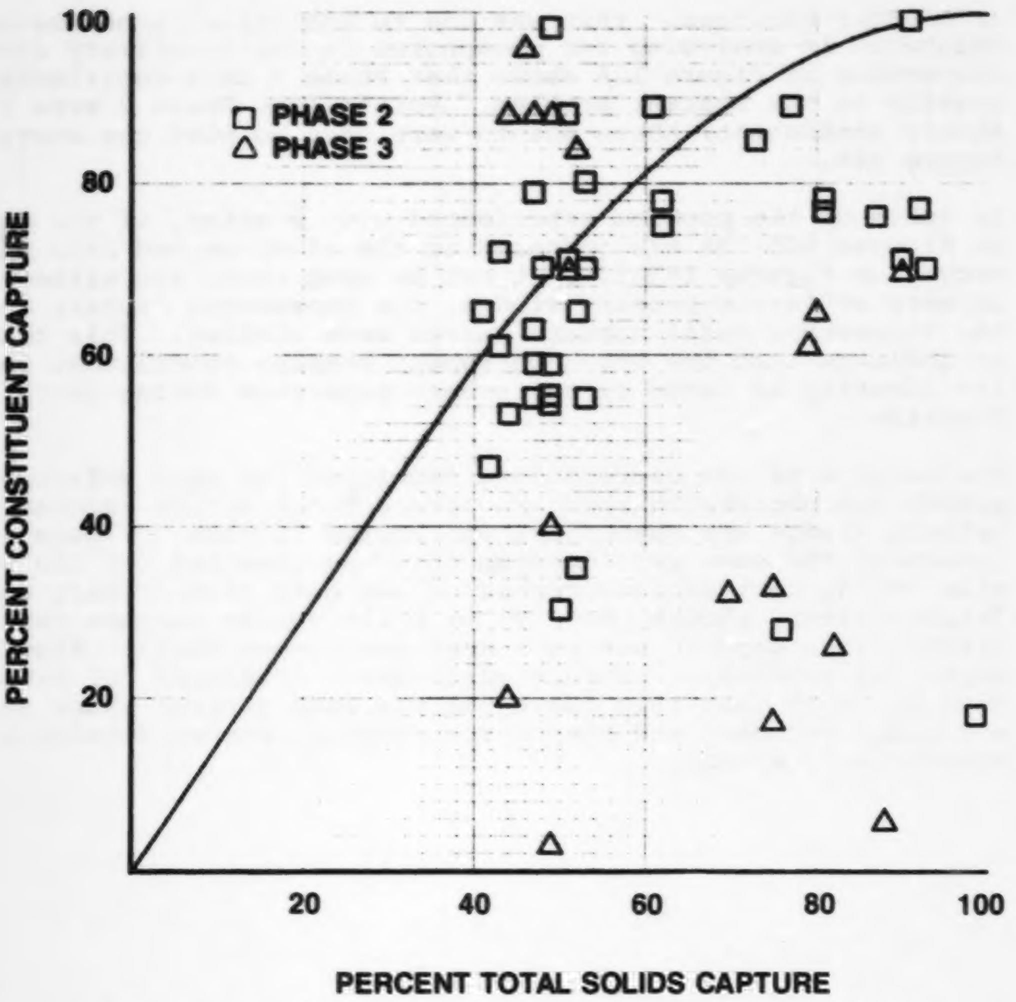

FIGURE 106. PHOSPHORUS CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF FERRIC-PRIMARY SLUDGE.

of scatter occurred. This was due to analytical problems encountered in analyzing for phosphorus in ferric-primary sludge. Inspection of Figure 106 shows that Phase 3 data contributed heavily to the scatter problem. Points from Phase 2 were relatively consistent; these points were used to plot the curve on Figure 106.

In spite of the problem experienced with scatter, if the curves on Figures 105-106 are compared to the aluminum and iron capture curves on Figures 103-104, it can be seen that, for either alum-primary or ferric-primary sludge, the phosphorus capture and the respective metal capture curves were similar. This tends to indicate that the metal-phosphate complex constituent retained its identity in terms of capture or rejection during centrifugation.

The results of the correlations developed for both volatile solids and phosphorus recovery versus total solids recovery for primary sludge are summarized on Figures 107-108. Figure 107 indicated the same relationship that was observed for the volatile solids correlations developed for both alum-primary and ferric-primary sludge, namely, volatile solids capture versus total solids capture was on a near one-to-one basis. Figure 108 shows the phosphorus capture correlation developed for primary sludge. Note that this curve has the same general shape as the aluminum recovery and phosphorus recovery curves developed from alum-primary sludge.

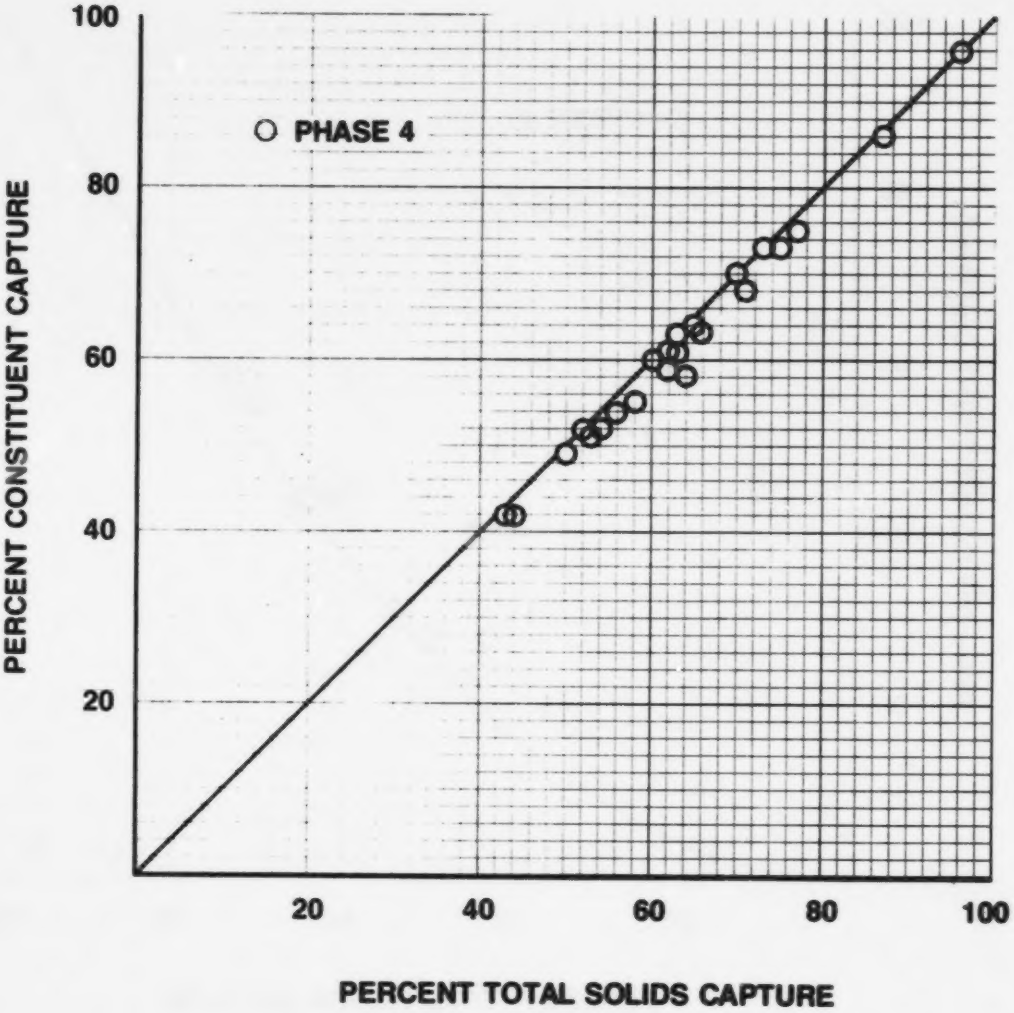

FIGURE 107. VOLATILE SOLIDS CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF PRIMARY SLUDGE.

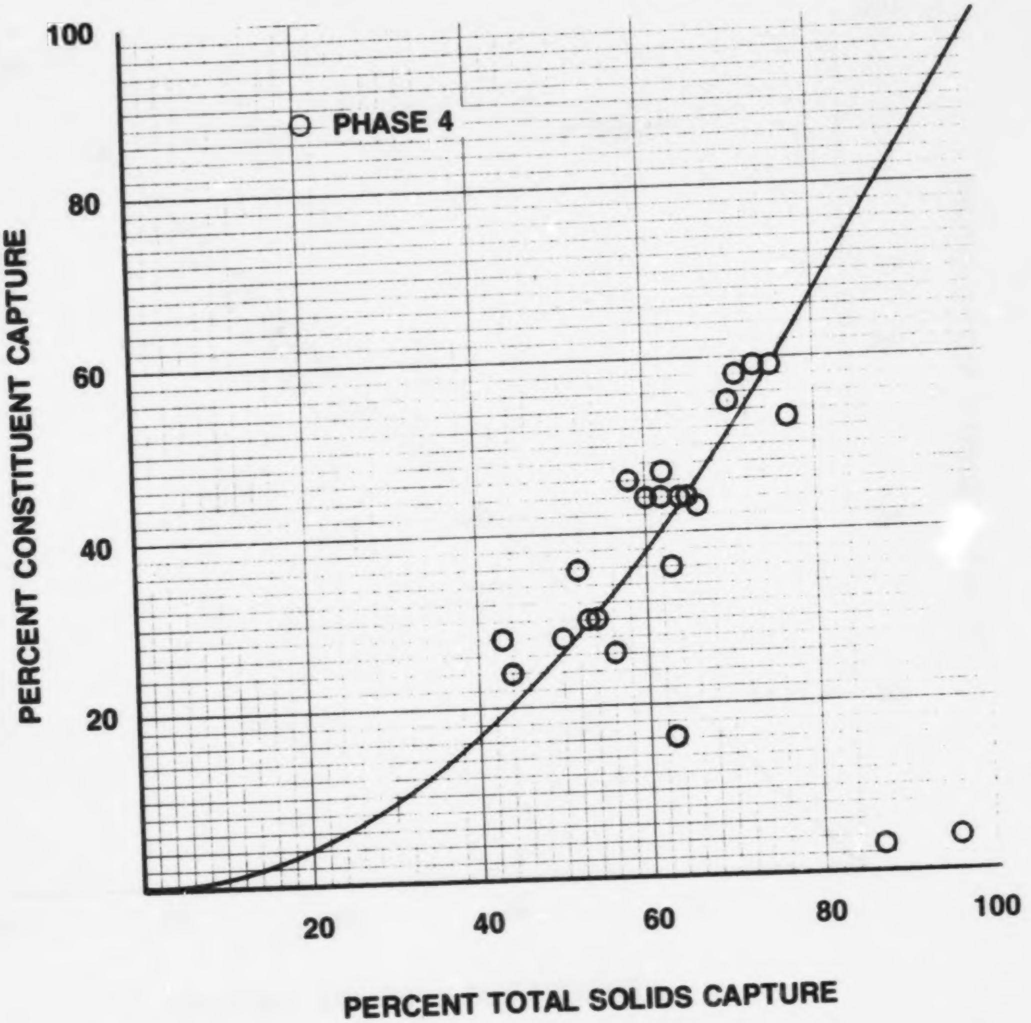

FIGURE 108- PHOSPHORUS CAPTURE VERSUS TOTAL SOLIDS CAPTURE FOR CENTRIFUGAL DEWATERING OF PRIMARY SLUDGE.

PRESSURE FILTRATION DEWATERING

Filter press studies commenced at the end of Phase 3 and then progressed into Phase 4. More meaningful data was generated in Phase 4 as Phase 3 proved to be primarily a shake-down period.

Results of all filter press test runs are presented in Table 47. An inspection of the results listed in Table 47 permit several generalized conclusions. For any specific sludge from Phase 4, filter press operation with thin filter cakes resulted in shorter press cycle times, higher filtration rates and cakes of greater solids concentration than resulted from press operation with thicker cakes.

Similar conclusions may be drawn from a review of the alum-primary sludge filter press results from Phase 3 operations. Shorter cycle times and higher filtration rates resulted from pressure filtration with thinner cakes. Generally, cake solids contents were equal for each cake thickness with the exception of run #10-24-74-1 where the thicker cake exhibited a higher cake solids than the thinner cake.

When comparing the three different types of sludges, some observations concerning performance can be made. When filter pressing primary sludge, the longest cycle times were required, but the cake solids were the lowest. Alum-primary sludge generated at a low alum dosage falls between the other two sludges, both in terms of cycle time and cake solids.

A filtration test in a filter press gives the time versus filtrate volume curve, and the dry solids yield and percent solids at the final filtration time. It is useful to be able to predict the moisture content of the cake if the form time were longer or shorter than the time selected. A predictive equation for cake moisture content can be developed from a dry solids and total solids balance around the filter press from time t to the final filtration time t_1:

Dry solids: $$D + (\Delta V)(C) = D_1 \qquad (1a)$$

Total solids: $$W + (\Delta V)(\Delta \rho') = W_1 \qquad (1b)$$

From Equations 1a and 1b, the cake solids content at time t can be calculated:

TABLE 47

SUMMARY OF PRESSURE FILTRATION RESULTS

Date Run No.	Type of Sludge	Unconditioned Feed Conc. (g/ℓ)	Conditioned Feed Conc. (g/ℓ)	Lime Conditioning (wt. %)	Cake Thickness		Cycle Time[3,4]		Cake Solids[6]		Cake Area		Filtration[7] Rate	
					Front (cm)	Back (cm)	Front (min)	Back (min)	Front (%)	Back (%)	Front (sq cm)	Back (sq cm)	Front (Kg/hr-sq m)	Back (kg/hr-sq m)
12-3-74-2	Alum[1]	33.5	41.8	37	3.5	2.7	104	87	26	27	455	494	.020	.023
12-3-74-3	Alum[1]	32.1	38.9	26	3.5	2.7	352	302	29	33	455	494	.007	.008
12-3-74-4	Alum[1]	24.3	32.8	36	3.5	2.7	262[5]	197	28	28	455	494	.008	.010
12-4-74-5	Alum[1]	33.0	35.2	33	3.5	2.7	142	116	26	26	455	494	.017	.022
11-19-74-1	Primary	57.3	59.7	37	3.5	2.7	369	222	34	42	455	494	.009	.014
11-10-74-2	Primary	57.3	66.7	44	3.5	2.7	340	266	25	33	455	494	.006	.010
11-21-74-3	Primary	28.9	49.2	48	3.5	2.7	650[5]	365	41	41	455	494	.003	.007
11-22-74-4	Primary	28.9	51.2	65	3.5	2.7	491[5]	415[5]	43	38	455	494	.005	.005
10-23-74-1	Alum[2]	49.7	54.5	30	2.9	2.7	302[5]	259[5]	33	33	506	494	.009	.010
10-23-74-2	Alum[2]	49.7	59.1	45	2.9	2.7	149	108	35	34	506	494	.021	.026
10-24-74-1	Alum[2]	30.2	39.7	33	2.9	2.7	494[5]	266	37	34	506	494	.005	.009
10-24-74-2	Alum[2]	30.2	40.5	45	2.9	2.7	223[5]	179	37	36	506	494	.012	.014

[1] Alum dosage in flocculating clarifier was 15.6 mg Al/ℓ

[2] Alum dosage in flocculating clarifier was 8.1 mg Al/ℓ

[3] Cycle time includes 20 minutes for press down-time during discharge and reloading

[4] All cycle times at 1/20 five minute rate

[5] This cycle time never reached, thie time extrapolated from filtrate volume versus time cuve

[6] Cake solids recalculated for all estimated cycle times

[7] Filtration rate for sludge solids only; conditioning chemicals subtracted out

Solids content (%) $$S_c = \frac{100D}{W} = \frac{D_1 - (\Delta V)(C)}{W_1 - (\Delta V)(\Delta \rho)} \qquad (1c)$$

where V = volume of filtrate at time t

S_c = percent dry solids in the cake

D = total solids in frames after V has been collected

W = total mass in frame after V has been collected

D_1 = total solids at end of filtration (V_1 has been collected)

W_1 = total mass at end of filtration (V_1 has been collected)

ΔV = final filtrate volume (V_1) minus filtrate volume at form time t (V)

C = conditioned feed concentration)g/l)

$\Delta\rho$ = difference in feed and filtrate densities (g/l)

The above equation is not rigorous unless concentrations and densities are based on final filtrate volume. However, it gives good approximations near the final filtering times. If S_c at t_1, D_1, W_1, V_1, $\Delta\rho$, and C are known, the value of S_c at any V can be found. If t versus V has been recorded in the course of the filtration test, then S_c versus t may be predicted.

The equation may be used to predict the cake solids (S_c) versus V relationship that would have been obtained had the press been allowed to run longer than the actual termination time of the test. The authors' experience indicates that this extrapolation should not be made if the filtration rate at the end of the test has dropped to below 1/20 of the average rate during the first five minutes. To predict the relationship between S_c and t for times greater than t_1, the t versus V curve must first be extrapolated.

Figures 109 illustrates the use of the equation for a primary sludge filter press run on 11-20-74. Note how the cake solids increased very little once the form time pertinent to 1/20 of the five minute rate had been reached.

If the filtration rate were calculated for different cake dry solids on the same filter press run, curves similar to those

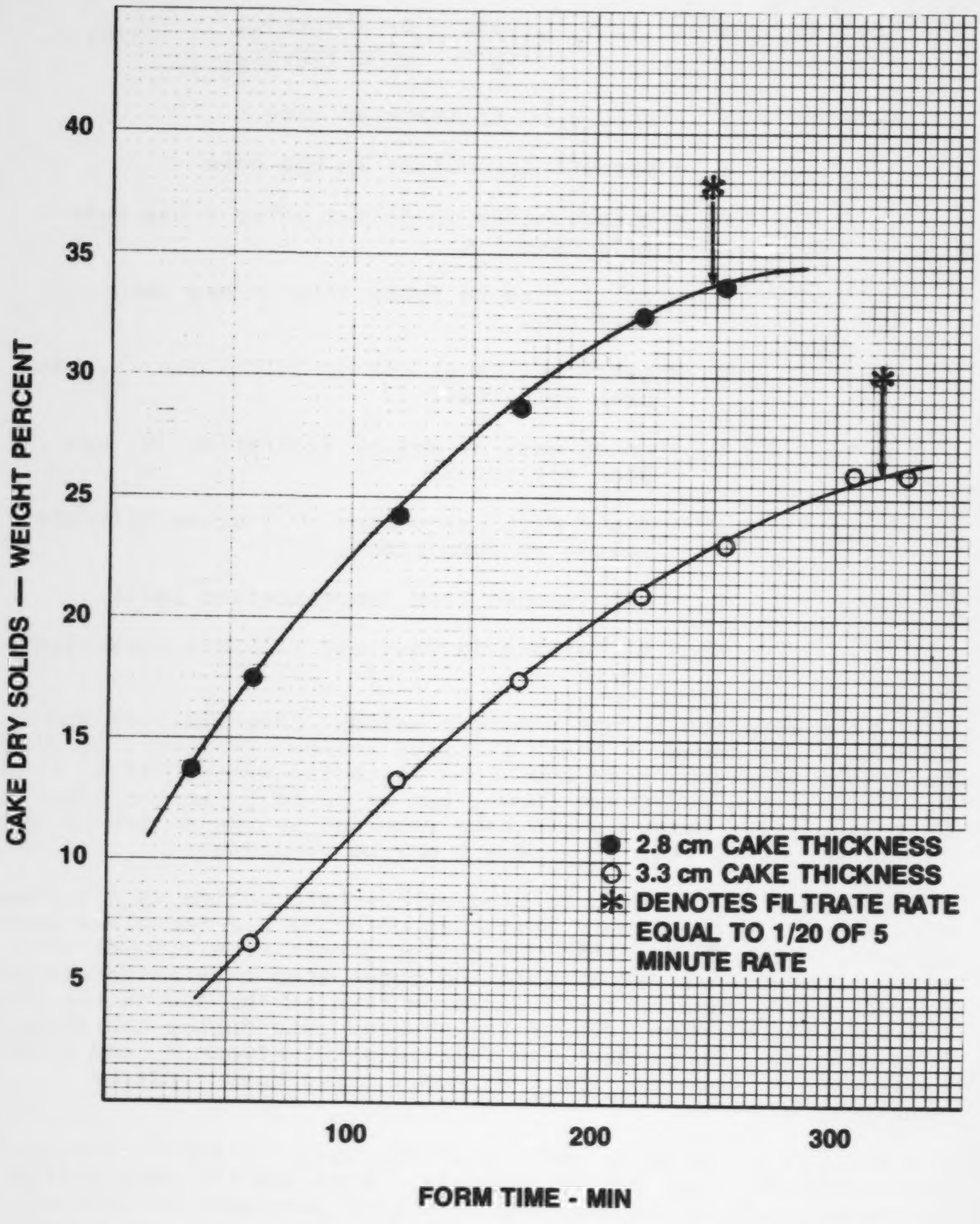

FIGURE 109. PRESSURE FILTER FORM TIME VS. CAKE SOLIDS CONCENTRATION FOR PRIMARY SLUDGE.

shown on Figures 110 through 112 would be generated. On Figure 110, two different filter press runs on alum-primary sludge are illustrated. In all cases, note the very rapid drop in filtration rate once 1/20 of the five minute rate had been reached; at this point and beyond, cake solids increased only slightly for a large loss in filtration rate. Figure 110 also shows the effect of cake thickness and lime dosage. Decreasing cake thickness and increasing lime dosage improved filtration rate.

The curves shown on Figures 110-112 can also be used to assist in selection of an operating point. By examining the curve on Figure 110, for a cake thickness of 3.5 cm (1.4 inch) and a feed concentration of 33 g/ℓ, it can be seen how drastically the filtration rate decreased after 25 weight percent cake solids was attained, and that 30 weight percent cake solids was unattainable.

Figures 111 and 112 generally showed the same pattern. Once 1/20 of the five minute filtrate production rate had been attained, filtration rate dropped markedly; also, thinner cakes provided higher filtration rates on a given sludge. On Figure 112, results for press runs on primary sludge were plotted. Results were consistent for cake thickness, however, results were inconsistent for level of lime conditioning. In this case, an increase of 7 percentage points in lime conditioning dose caused a drop in the filtration rate (for the same feed solids). Since the run using 44 weight percent lime was not conducted the same day as the one with 37 weight percent lime, the reason for the decrease in filtration rate could be attributed to different sludge characteristics. It is felt that the test with 44 percent lime conditioning was an overdose (for example, Figure 8 shows that any lime conditioning in excess of 30 weight percent on that sludge was of no value in further reduction of specific resistance).

Some general observations can be made concerning pressure filtration of alum-primary sludge and primary sludge. When pressing alum-primary sludge (generated at both a high and low alum dosage) lime conditioning in the range of 24 to 25 percent by weight was required; however, for primary sludge a higher range of 37 to 65 percent was required. On alum-primary sludge (high dosage), a decrease in the cake thickness from 3.5 to 2.7 cm (1.4-1.1 inch) resulted in a filtration rate increase of from 14 to 29 percent; cake solids generally increased about 2 percentage points for this cake thickness decrease. For alum-primary sludge generated at a low alum dosage, a much smaller cake thickness difference (2.9 cm to 2.7 cm {1.1-1.0 inch}) generally resulted in filtration rate increases of from 11 to 24 percent, while cake solids showed little difference. On primary sludge, the cake thickness difference had the most significance. A cake thickness decrease from 3.5-2.7 cm (1.4-1.1 inch) resulted in filtration rate increases of from 56 to greater than 100 percent, while cake solids increased as much as 8 percentage points for this cake

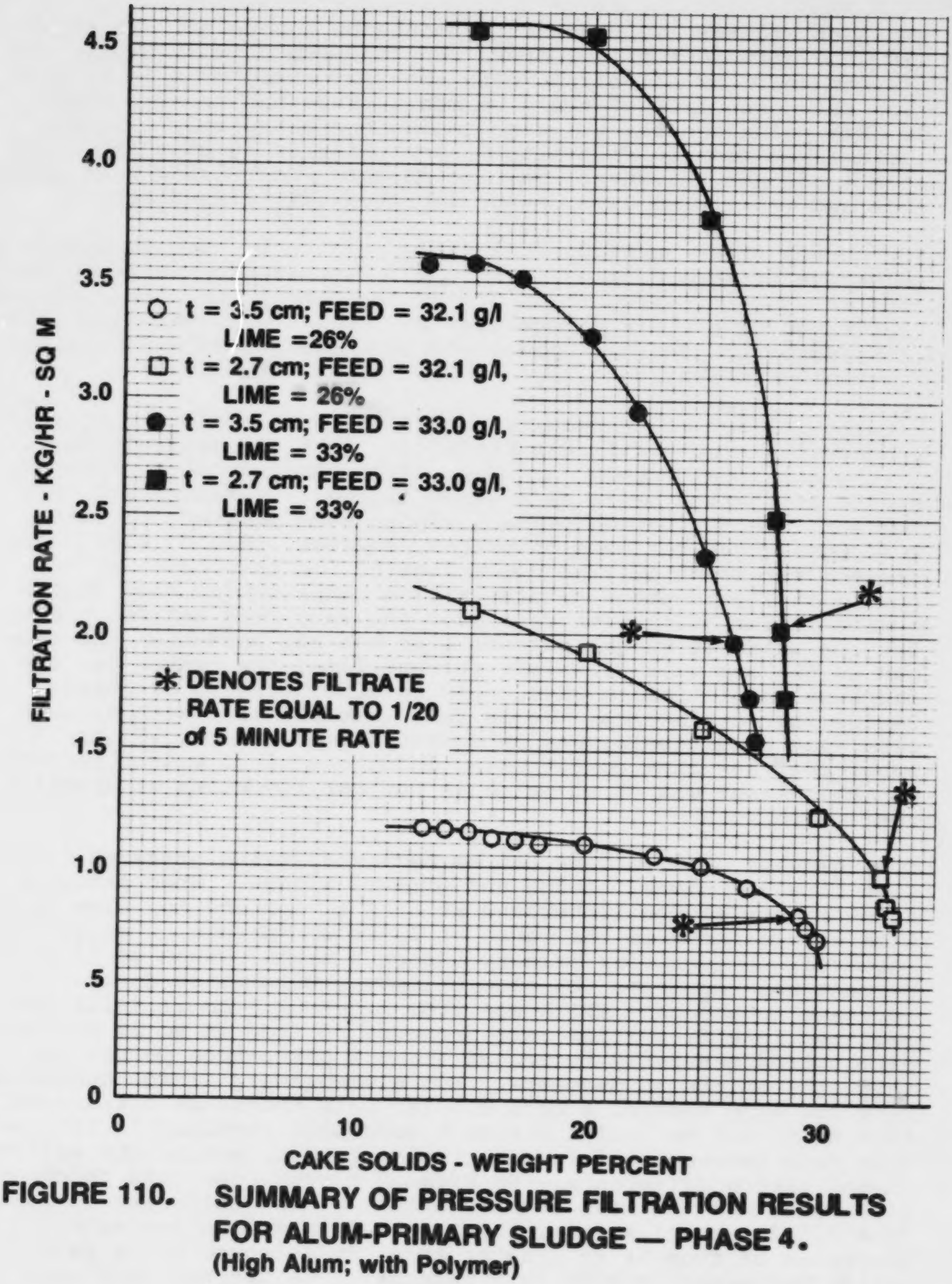

FIGURE 110. SUMMARY OF PRESSURE FILTRATION RESULTS FOR ALUM-PRIMARY SLUDGE — PHASE 4.
(High Alum; with Polymer)

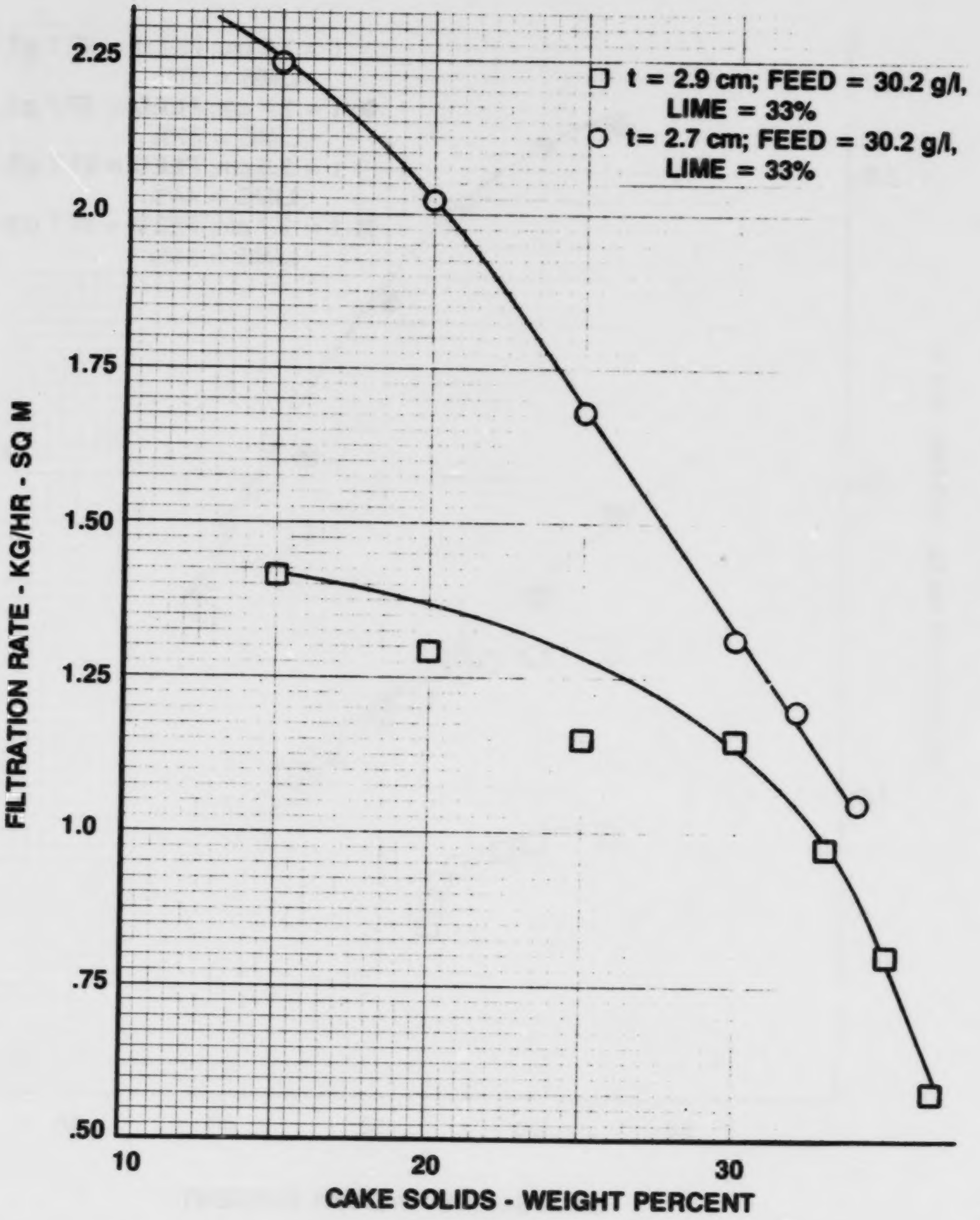

FIGURE 111. SUMMARY OF PRESSURE FILTRATION RESULTS FOR ALUM-PRIMARY SLUDGE - PHASE 3. (Low Alum; with Polymer)

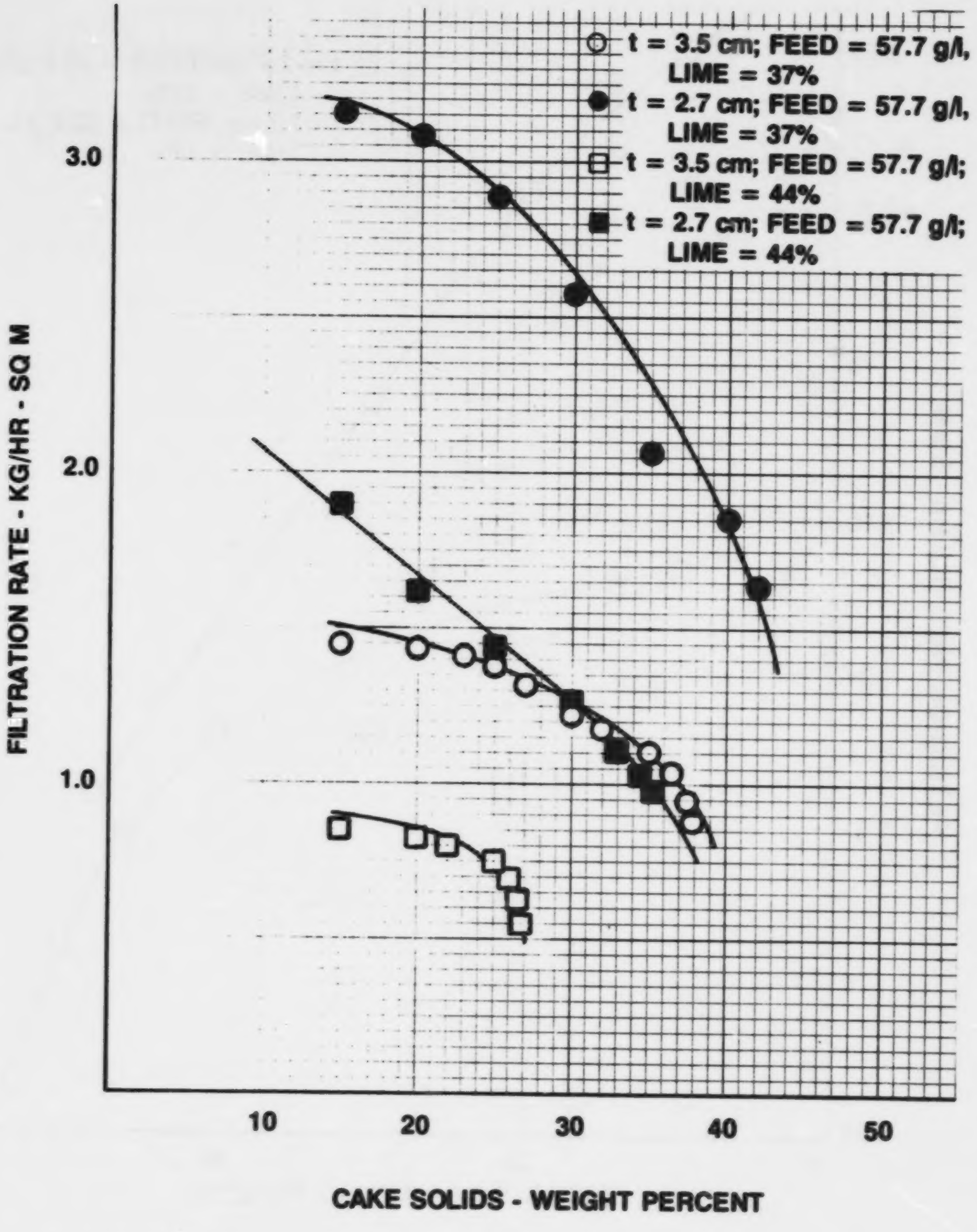

FIGURE 112. SUMMARY OF PRESSURE FILTRATION RESULTS FOR PRIMARY SLUDGE.

thickness decrease. Filtration rates were lowest for primary sludge, generally about one-half the rates for alum-primary sludge generated at either alum dosage. Cake solids were higher for primary sludge, generally 8 to 9 percentage points above alum-primary sludge generated at a high alum dosage.

SECTION VII

DEWATERING PERFORMANCE AND COST COMPARISON

One of the key objectives of this study was to compare the application of solid bowl centrifugation and vacuum belt filtration to the dewatering of chemical-primary sludges generated under various pilot plant operating modes. The purpose of this Section is to present a comparative summary of performance achieved and to establish the relative costs for these two dewatering techniques for specific chemical-primary sludges.

The approach used to develop the comparative performance and costs for these two dewatering techniques involved:

1. The selection of a municipal wastewater with typical concentrations of BOD_5, TSS and phosphorus and,

2. a prediction of the characteristics of the sludges produced from chemical treatment of this wastewater using alum or ferric chloride, each at two specific levels of phosphorus removal.

The prediction of sludge characteristics was formulated based on the various analyses made during this study. To review the relative costs of the two dewatering approaches, plant sizes of 20,000, 40,000 and 200,000 cu m/day were selected. A detailed summary of assumptions made for this performance and cost analysis is contained in Appendix G.

Table 48 describes the predicted sludge characteristics that results from chemical treatment of a typical municipal wastewater using either alum or ferric chloride to remove phosphorus.

From the various analyses made during this study, the dewatering characteristics of the sludges described in Table 48 were predicted. These predictions were used to compare performance between dewatering methods, chemical coagulant and degree of phosphorus removal. Table 49 contains a summary of comparative dewatering performance results.

The results in Table 49 indicates that dewatering ferric-primary sludge would require more chemical conditioning and more machine capacity than that for alum-primary sludge. Cake solids contents produced from either method were significantly higher when

TABLE 48

SUMMARY OF SLUDGE CHARACTERIZATION FROM CHEMICAL TREATMENT OF MUNICIPAL WASTEWATER

Sludge Characterization	Percent P Removal with Primary Coagulant			
	Alum		Ferric Chloride	
	80	95	80	95
Sludge production, Kg DS/cu m	0.221	0.247	0.228	0.258
Volatiles, wt. % of total	69.1	68.8	67.2	59.2
Inerts, wt. % of total	12.2	10.9	11.8	10.4
Chemical, wt. % of total	18.7	27.3	21.0	30.4
Thickened feed sludge to dewatering, wt. % TS	4.6	3.8	5.6	5.4

TABLE 49

SUMMARY OF DEWATERING PERFORMANCE FOR CHEMICAL-PRIMARY SLUDGES

Dewatering Performance	Percent P Removal with Primary Coagulant			
	Alum		Ferric Chloride	
	80	90	80	90
Vacuum Filtration				
1. Chemical conditioning lime dose, %	25	25	30	30
2. Filter yield, Kg/hr-sq m(2)	13.5	8.6	10.2	8.6
3. Cake				
Solids content, % TS(1)	25.5	25.6	34.5	34.5
Volatile content, %(1)	55	49	52	46
4. Solids capture, %	97-99	97-99	97-99	97-99
Centrifuguation				
1. Chemical conditioning polymer dose, %	0.3	0.3	0.35	0.35
2. Loading, ℓpm(3)	8.7	8.7	6.7	6.7
3. Cake				
Solids content, % TS(1)	18	18	22	22
Volatile content, %(1)	69	62	69	62
4. Solids capture, %	95	95	90	90

(1) Including chemicals

(2) Excluding chemicals

(3) Solid bowl at 15.2 cm x 31.4 cm

dewatering ferric-primary sludge as compared to alum-primary sludge.

The results summarized in Table 49 also indicate that the solids content of the filter cakes produced by vacuum filtration of either chemical-primary sludge were substantially greater than those produced by centrifugation. Cake volatility, however, was lower for the filter cakes due to lime conditioning required for filtration dewatering as compared to polymer conditioning required for centrifugal dewatering.

An inspection of the results summarized in Table 49 indicates that vacuum filtration rates for either chemical-primary sludge produced with 95 percent phosphorus removal were lower than for sludges produced with 80 percent phosphorus removal. Filter cake solids contents were not significantly affected by level of phosphorus removal between 80 to 95 percent. Deterioration in machine capacity was not observed for centrifugal dewatering. Generally, operation at 95 percent phosphorus removal as compared to 80 percent removal resulted in a greater sludge production and in a cake of lower volatility.

The results given in Tables 48 and 49 were used to establish cost information for vacuum filtration and centrifugal dewatering alternatives for specific chemical-primary sludges.

The results of the cost analysis are summarized in Tables 50-53. As shown, the cost analysis was performed for two levels of phosphorus removal. This was done in order to study the implications of the level of phosphorus removal attained from chemical treatment of raw wastewater on the costs of sludge disposal. Many of the analyses developed from the results obtained during this pilot plant study indicated that the thickening and dewatering characteristics of chemical-primary sludges deteriorated as the fraction of chemical solids contained in the sludge increased. It was reasoned, therefore, that the sludge disposal costs associated with 95 percent phosphorus removal would be higher than the sludge disposal costs associated with 80 percent phosphorus removal.

Estimated incineration costs were developed in addition to dewatering costs. This was done in order to account for the variation in the characteristics (moisture content, volatile content) of the cakes produced by the dewatering machines.

Inspection of the annual cost data shown in Tables 50-53 indicated the following conclusions:

TABLE 50

COST SUMMARY OF VACUUM FILTRATION OF ALUM-PRIMARY SLUDGE

Cost	80 Percent P Removal			95 Percent P Removal		
	Plant Size - cu/m day			Plant Size - cu m/day		
1000 Dollars	20,000	40,000	200,000	20,000	40,000	200,000
DEWATERING						
Capital cost	270	266	664	389	351	1087
Annual power cost	5	9	28	8	13	52
Annual labor cost	8	32	63	8	31	62
Annual chemical cost	20	40	200	22	45	223
Annual maintenance cost	4	4	14	7	7	26
Annual amortized cost	28	27	68	40	36	111
TOTAL ANNUAL COST	65	112	373	85	132	474
TOTAL COST, Dollars/Metric TDS	40	35	23	47	37	26
INCINERATION						
Capital cost	1512	1512	3024	1512	1512	3938
Annual O&M cost	76	81	248	80	90	317
Annual amortized cost	154	154	308	154	154	401
TOTAL ANNUAL COST	230	235	556	234	244	718
TOTAL COST, Dollars/Metric TDS	143	73	34	130	68	40
TOTAL DISPOSAL COST						
Dollars/Metric TDS	183	108	57	177	105	66

TABLE 51

COST SUMMARY OF VACUUM FILTRATION OF FERRIC-PRIMARY SLUDGE

Cost 1000 Dollars	80 Percent P Removal			95 Percent P Removal		
	Plant Size - cu m/day			Plant Size - cu m/day		
	20,000	40,000	200,000	20,000	40,000	200,000
DEWATERING						
Capital cost	313	306	809	388	339	1088
Annual power cost	6	11	37	8	15	54
Annual labor cost	8	32	65	8	30	64
Annual chemical cost	25	49	247	28	56	208
Annual maintenance cost	5	4	18	7	6	26
Annual amortized cost	32	31	82	40	35	111
TOTAL ANNUAL COST	76	127	449	91	142	535
TOTAL COST, Dollars/Metric TDS	46	38	27	48	38	28
INCINERATION						
Capital cost	1096	1096	2520	1260	1298	2646
Annual O&M cost	45	38	98	49	46	102
Annual amortized cost	112	112	257	128	132	270
TOTAL ANNUAL COST	157	150	355	177	178	372
TOTAL COST, Dollars/Metric TDS	95	45	21	94	47	20
TOTAL DISPOSAL COST						
Dollars/Metric TDS	141	83	48	142	85	48

TABLE 52

COST SUMMARY OF CENTRIFUGATION OF ALUM-PRIMARY SLUDGE

Cost	80 Percent P Removal			95 Percent P Removal		
	Plant Size - cu m/day			Plant Size - cu m/day		
1000 Dollars	20,000	40,000	200,000	20,000	40,000	200,000
DEWATERING						
Capital cost	276	276	686	343	343	852
Annual power cost	7	13	40	8	14	54
Annual labor cost	9	34	68	8	28	68
Annual chemical cost	22	45	224	26	53	264
Annual maintenance cost	6	6	21	7	7	28
Annual amortized cost	28	28	70	35	35	87
TOTAL ANNUAL COST	72	126	423	84	137	501
TOTAL COST, Dollars/Metric TDS	45	39	26	47	38	28
INCINERATION						
Capital cost	1512	1575	4095	1890	1890	4253
Annual O&M cost	91	116	441	117	150	551
Annual amortized cost	154	160	417	193	193	443
TOTAL ANNUAL COST	245	276	858	310	343	984
TOTAL COST, Dollars/Metric TDS	152	86	53	172	95	55
TOTAL DISPOSAL COST						
Dollars/Metric TDS	197	125	79	219	133	83

TABLE 53

COST SUMMARY OF CENTRIFUGATION OF
FERRIC-PRIMARY SLUDGE

Cost 1000 Dollars	80 Percent P Removal Plant Size - cu m/day 20,000	 40,000	 200,000	95 Percent P Removal Plant Size - cu m/day 20,000	 40,000	 200,000
DEWATERING						
Capital cost	339	339	842	339	339	844
Annual power cost	7	11	45	8	13	53
Annual labor cost	7	24	59	8	28	68
Annual chemical cost	28	56	281	32	63	316
Annual maintenance cost	7	7	28	7	7	28
Annual amortized cost	35	35	86	35	35	86
TOTAL ANNUAL COST	84	133	499	90	146	551
TOTAL COST, Dollars/Metric TDS	51	40	30	48	39	29
INCINERATION						
Capital cost	1575	1827	3024	1575	1575	3938
Annual O&M cost	79	90	221	87	100	326
Annual amortized cost	160	186	308	160	160	401
TOTAL ANNUAL COST	239	276	529	247	260	727
TOTAL COST, Dollars/Metric TDS	144	83	32	131	69	39
TOTAL DISPOSAL COST						
Dollars/Metric TDS	195	123	62	179	108	68

1. It was costlier by 10-20 percent to centrifugally dewater and 15-25 percent to vacuum filter dewater alum-primary or ferric-primary sludge produced from chemical treatment aimed at 95 percent phosphorus removal as compared to 80 percent phosphorus removal. This was caused by the increased quantities of sludge and a generally more difficult sludge to dewater for the 95 percent phosphorus removal case.

2. There was no significant difference between centrifugal and vacuum filtration dewatering costs for either alum-primary or ferric-primary sludge produced from chemical treatment aimed at 95 percent phosphorus removal. The total costs (dewatering plus cake incineration) were 25-30 percent higher when centrifugation was used as compared to vacuum filtration.

3. Centrifugal dewatering was 10-15 percent more expensive than vacuum filtration of either alum-primary or ferric-primary sludge produced from chemical treatment aimed at 80 percent phosphorus removal. The total costs (dewatering plus cake incineration) were 10-35 percent higher when centrifugation was used as compared to vacuum filtration for the alum-primary sludge case. The total costs (dewatering plus cake incineration) were 30-45 percent higher when centrifugation was used as compared to vacuum filtration for the ferric-primary sludge case.

The conclusions listed above indicate the increased sludge handling costs realized as a result of chemical treatment of municipal wastewater to effect 95 percent phosphorus removal as compared to lower levels of phosphorus removal, i.e., 80 percent. The reasons for the increased cost are the deteriorated sludge dewatering properties which are documented in this report and the increased quantities of waste sludge generated with operation at 95 percent phosphorus removal.

The conclusions also indicate that centrifugal dewatering was generally costlier than vacuum filtration dewatering particularly when the cost of cake disposal (i.e., incineration) was considered in the economic analysis.

Figures 113-116 summarize the total dewatering costs and total disposal costs (including incineration) on a unit cost basis (dollars per metric ton of dry solids) for the various process and machine options studied. When viewed on the basis of unit costs, the results on Figures 113-116 indicate the following conclusions:

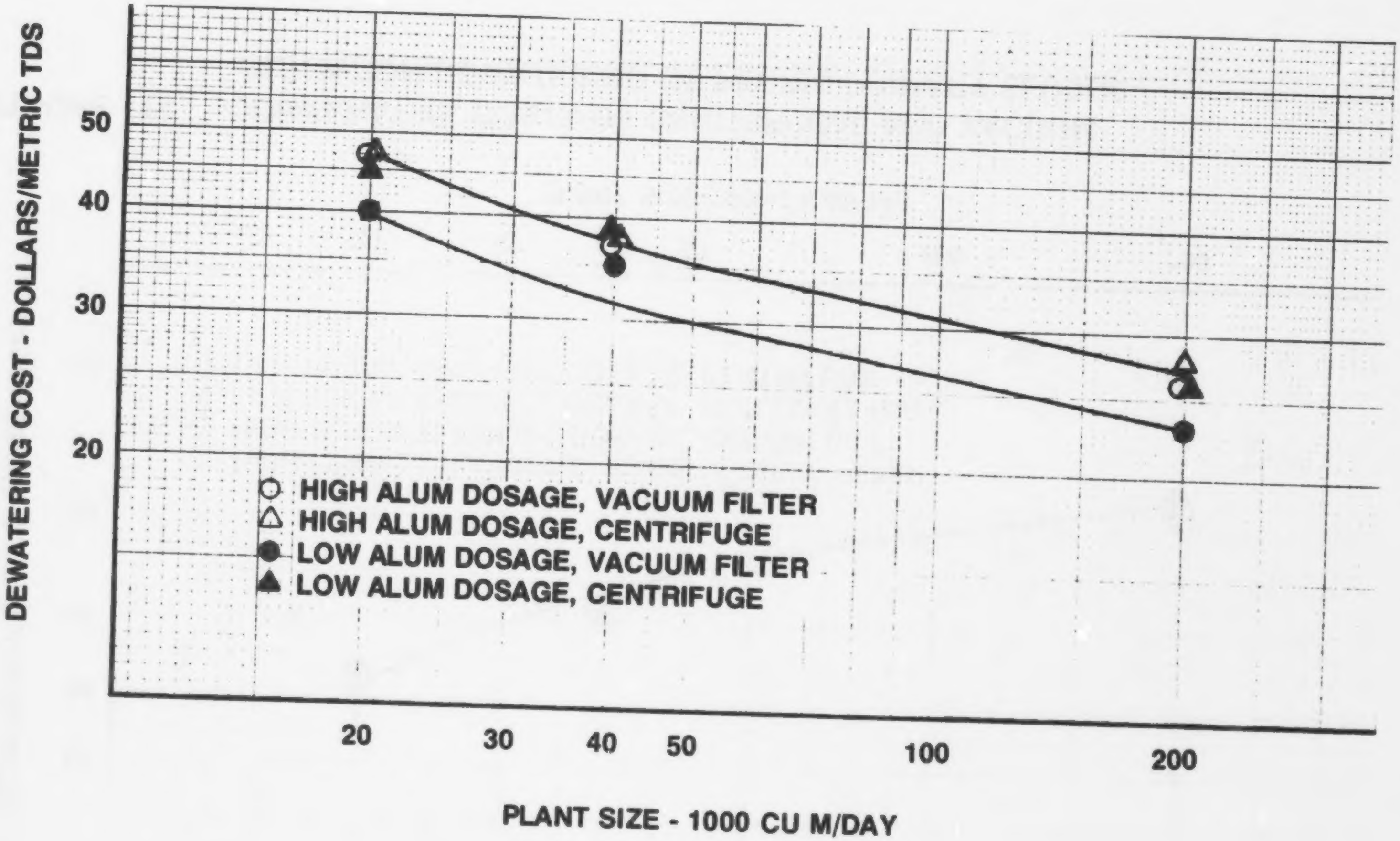

FIGURE 113. COMPARATIVE COSTS OF CENTRIFUGAL AND VACUUM FILTRATION DEWATERING OF ALUM-PRIMARY SLUDGE.

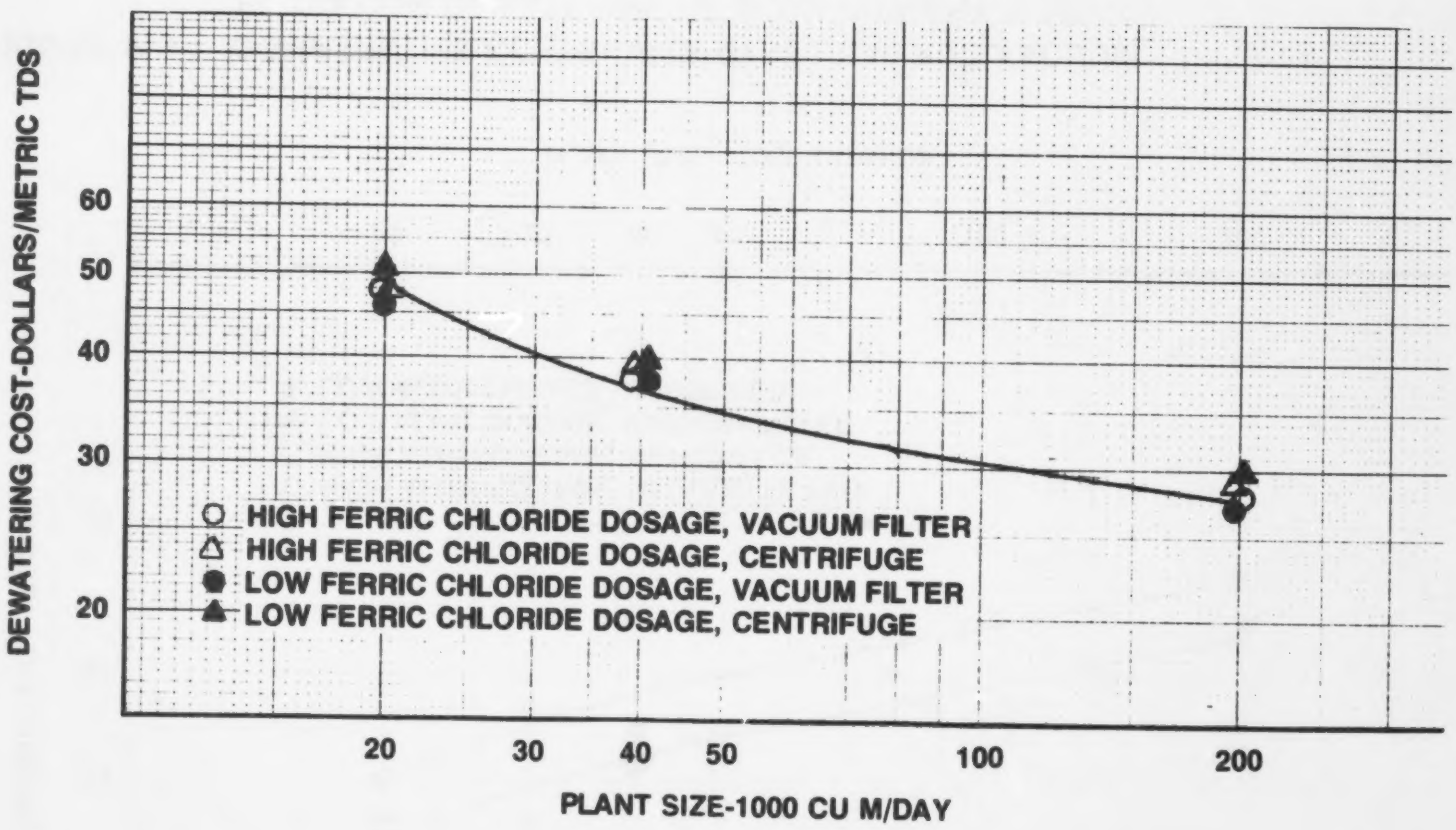

FIGURE 114. COMPARATIVE COSTS OF CENTRIFUGAL AND VACUUM FILTRATION DEWATERING OF FERRIC-PRIMARY SLUDGE.

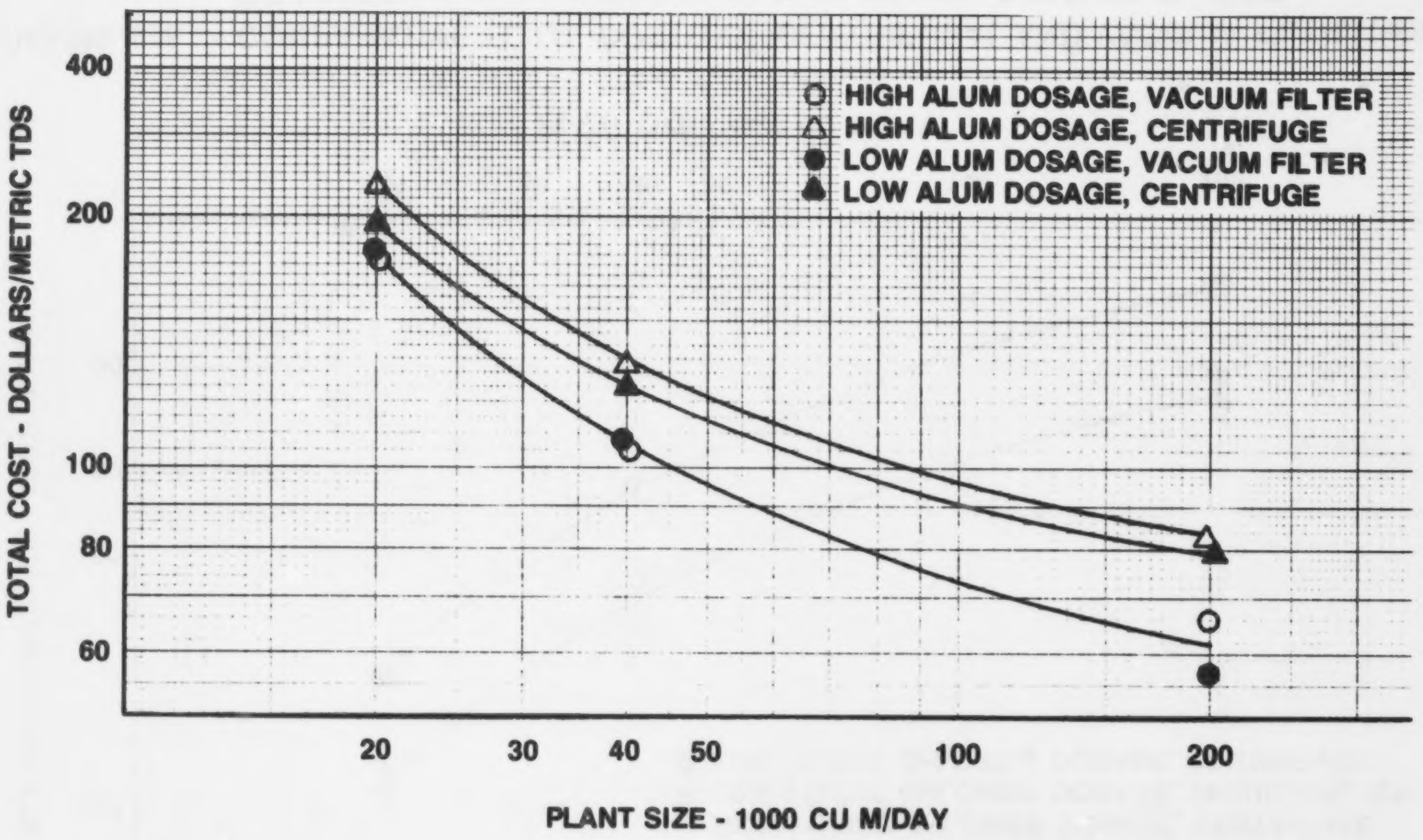

FIGURE 115. COMPARATIVE TOTAL COSTS OF CENTRIFUGAL AND VACUUM FILTRATION DEWATERING PLUS INCINERATION OF ALUM-PRIMARY SLUDGE.

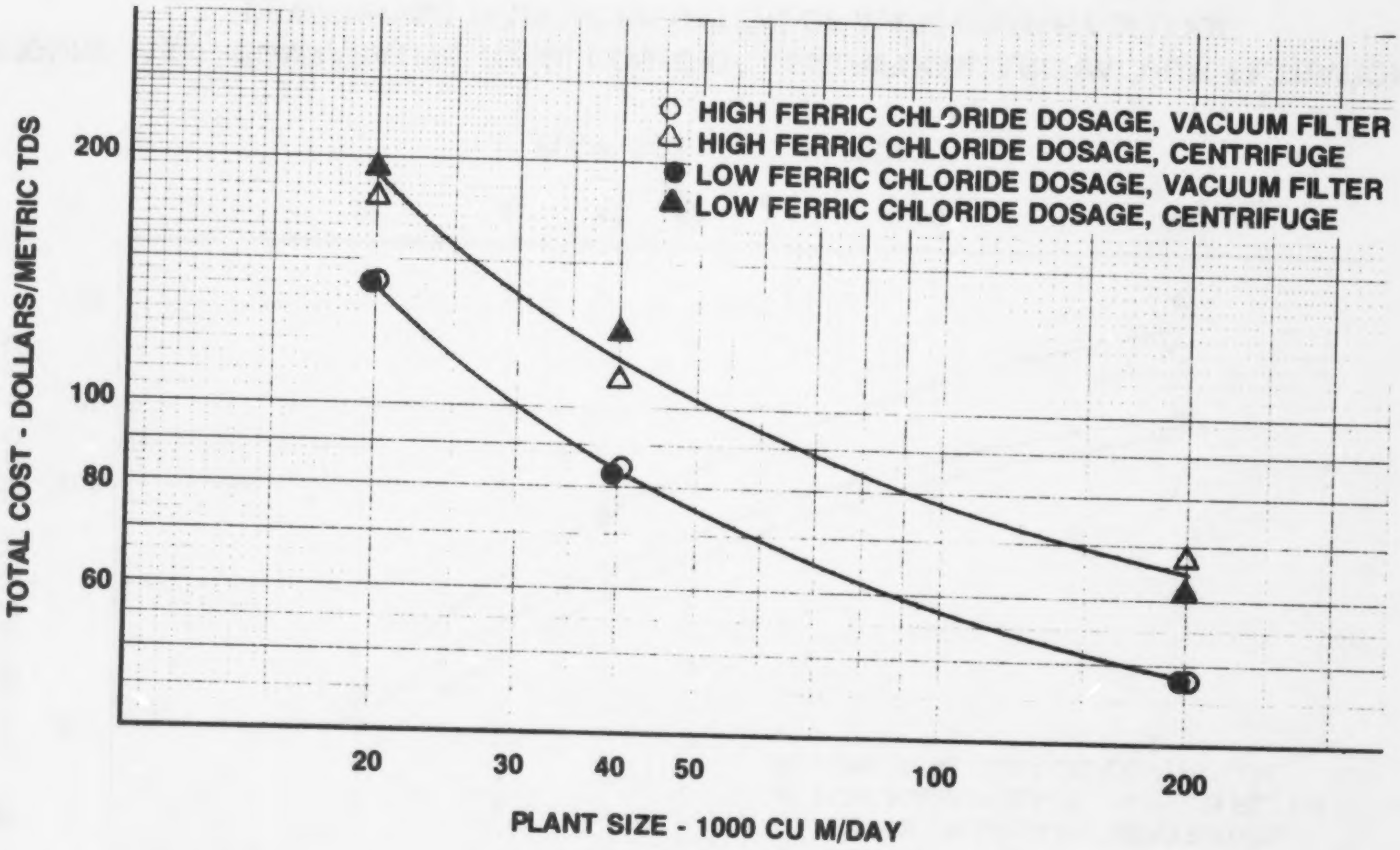

FIGURE 116 — COMPARATIVE TOTAL COSTS OF CENTRIFUGAL AND VACUUM FILTRATION DEWATERING PLUS INCINERATION OF FERRIC-PRIMARY SLUDGE.

1. There was no significant difference in dewatering costs for centrifugal dewatering of sludge produced from chemical treatment aimed at phosphorus removal of 80 percent versus 95 percent. This conclusion was indicated whether alum or ferric chloride was used as the primary coagulant.

2. Alum-primary sludge produced from 95 percent phosphorus removal was 10-15 percent more costly to vacuum filter dewater than alum-primary sludge produced from 80 percent phosphorus removal. There was no significant difference in these costs for the case of ferric-primary sludges.

3. There was no significant difference between centrifugal and vacuum filtration dewatering costs for alum-primary sludge produced from 95 percent phosphorus removal. The same was the case for ferric-primary sludge produced from 95 percent phosphorus removal.

4. Centrifugal dewatering was 10-15 percent more expensive than vacuum filtration of alum-primary sludge produced from 80 percent phosphorus removal. When dewatering ferric-primary sludge produced from 80 percent phosphorus removal, the centrifugal dewatering approach was approximately 10 percent more costly than the vacuum filtration approach.

5. There was no significant difference in total costs (dewatering plus cake incineration) for centrifugal dewatering of sludge produced from 80 percent versus 95 percent phosphorus removal. This conclusion was indicated whether alum or ferric chloride was used as the primary coagulant.

6. There was no significant difference in total costs (dewatering plus cake incineration) for vacuum filtration dewatering of sludge produced from 80 percent versus 95 percent phosphorus removal. This conclusion was indicated whether alum or ferric chloride was used as the primary coagulant.

7. The total costs (dewatering plus cake incineration) were approximately 25 percent higher when centrifugation was used as compared to vacuum filtration of alum-primary sludge produced from 95 percent phosphorus removal. The cost differential increased to approximately 30 percent for the case of ferric-primary sludge produced from 95 percent phosphorus removal.

8. The total costs (dewatering plus cake incineration) averaged approximately 20 percent higher when centrifugation was used as compared to vacuum filtration of alum-primary sludge produced from 80 percent phosphorus removal. The cost differential increased to approximately 40 percent for the case of ferric-primary sludge produced from 80 percent phosphorus removal.

9. Operating costs comprise 60-80 percent of the dewatering costs for either machine dewatering either chemical-primary sludges.

10. Operating costs comprise 40-60 percent of the total (dewatering plus cake incineration) costs for either machine dewatering either chemical-primary sludge.

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APPENDIX A

DISSOLVED AIR FLOTATION DATA REDUCTION PROCEDURE

Sample Calculation

1. Data collection

(a) Recycle liquid temperature, 20.0°C
(b) Flotation chamber calibration 60.0 mℓ/cm
(c) Initial slurry height, 0.167 m
(d) Feed concentration, 16.5 g/ℓ
(e) Feed volume, 400 mℓs
(f) Recycle volume, 600 mℓs
(g) Recycle pressure, 4.57 Kg/sq cm
(h) Barometric pressure, 0.896 Kg/sq cm

2. Calculations

(a) Initial slurry concentration

$$C_o = \frac{(16.5\ \mathrm{g}/\ell)(0.400\ell)}{(0.400\ell + 0.600\ell)} = 6.60\ \mathrm{g}/\ell$$

(b) Solids loading rate (SLR)

$$\mathrm{SLR} = \frac{(6.60\ \mathrm{g}/\ell)(1000\ \ell/\mathrm{cu\ m})(0.167\ \mathrm{m})(60\ \mathrm{min/hr})}{(4.1\ \mathrm{min})(1000\ \mathrm{g/Kg})}$$

$$\mathrm{SLR} = 16.1\ \mathrm{Kg/sq\ m\text{-}hr}$$

(c) Full-scale flotation rate (FSFR)

$$\mathrm{FSFR} = (0.667^{*})\ (16.1\ \mathrm{Kg/sq\ m\text{-}hr})$$

$$\mathrm{FSFR} = 10.8\ \mathrm{Kg/sq\ m\text{-}hr}$$

(d) Hydraulic loading rate (HLR)

(1) $$\mathrm{HLR} = \frac{(10.8\ \mathrm{Kg/sq\ m\text{-}hr})(1000\ \mathrm{g/Kg})}{(6.60\ \mathrm{g}/\ell)(1000\ \ell/\mathrm{cu\ m})}$$

$$\mathrm{HLR} = 1.64\ \mathrm{cu\ m/sq\ m\text{-}hr}\ \text{(includes recycle)}$$

*Scale-up factor

(2) Bulk separation rate (BSR)

$$\text{BSR} = \frac{(330\ \text{m}\ell/\text{min})(60\ \text{min/hr})(1000\ \ell/\text{cu m})(0.5^*)}{(60\ \text{m}\ell/\text{cm})(100\ \text{cm/m})(1000\ \ell/\text{cu m})}$$

$$\text{BSR} = 1.65\ \text{cu m/sq m-hr}$$

(e) Air to solids ratio (A/S)

$$\text{g air} = (0.129\ \text{g}/\ell)(0.600\ell) - (0.0211\ \text{g}/\ell)(0.600\ell) = 0.0647\ \text{g}$$

$$\text{g dry solids} = (16.5\ \text{g}/\ell)(0.400\ell) = 6.60\ \text{g}$$

$$\text{A/S} = 0.980\%$$

---

[*] Scale-up factor

WEIGHT OF AIR DISSOLVED IN WATER AT 4.57 Kg/sq cm

T, °C	S_g, G/L at 1 ATM
15	0.0266
16	0.0260
17	0.0255
18	0.0252
19	0.0247
20	0.0244
21	0.0240
22	0.0236
23	0.0233
24	0.0229
25	0.0226

$$\text{G/L of Air} = S_g \frac{190 + P}{43.0}$$

P = barometric pressure, Kg/sq cm

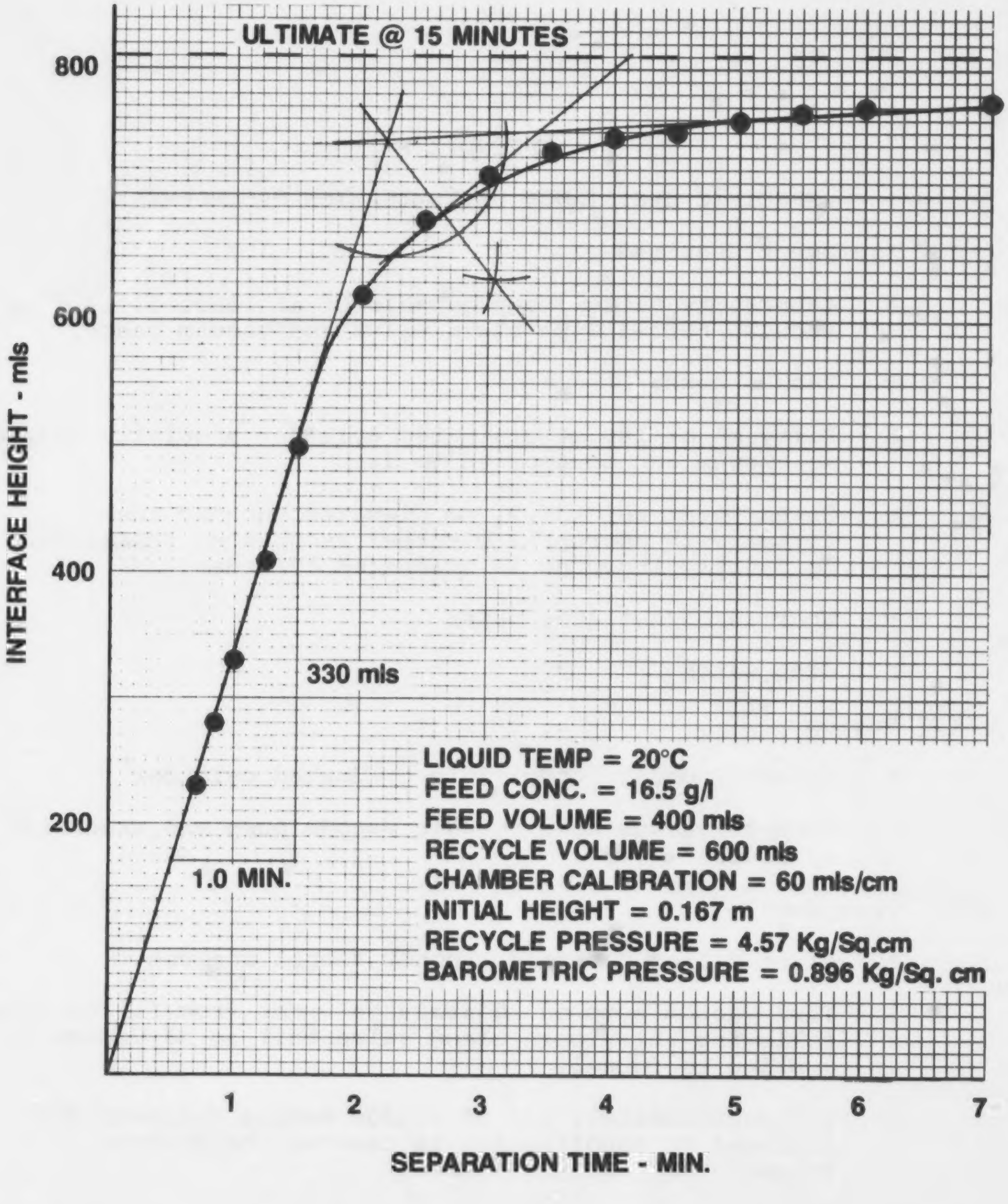

FIGURE A1.— SAMPLE PLOT OF TYPICAL DAF THICKENING DATA.

APPENDIX B

SPECIFIC RESISTANCE DETERMINATION PROCEDURES

I. Equipment Required

   A. Laboratory vacuum pump capable of at least 51 cm. Hg vacuum, fitted with bleed valve and vacuum gauge.

   B. Vacuum hose

   C. Fifty mℓ or 100 mℓ graduated cylinder specially adapted so vacuum can be applied to it.

   D. Seven point zero (7.0) cm diameter Buchner funnel fitted with appropriate rubber stopper so funnel can be tightly inserted in graduated cylinder.

   E. Whatman GF/A filter paper

   F. Stop watch

II. Set-up

   A. Insert Buchner funnel into graduated cylinder

   B. Connect vacuum hose between vacuum pump and graduated cylinder

III. Procedure

   A. Place filter paper in Buchner funnel and wet.

   B. Turn vacuum pump on to evacuate water from filter paper; following this, open bleed valve full so no vacuum is applied to funnel.

   C. Add approximately 200 mℓ sludge sample (already conditioned if conditioning is desired) to Buchner funnel.

   D. Quickly close bleed valve so vacuum of 50.8 cm Hg is applied to funnel.

   E. When desired vacuum level is reached, note filtration volume and start stop watch.

F. Record time and filtrate volume as test progresses. As this is done, a chart as shown below will be developed.

Time (sec)	Filtrate Volume (mℓ)	Corrected Filtrate Volume (Vc) (mℓ)	θ/V_1
0	V_o	$V_o - V_o$	-
θ_1	V_1	$V_1 - V_o$	θ_1/V_c
θ_2	V_2	$V_2 - V_o$	θ_2/V_c

Specific Resistance Calculation

$$R = \frac{2\ PA^2}{\mu\ c}\ b$$

where - R = specific resistance, sec^2/g
P = filtration pressure, g/sq cm
A = filter area, sq cm
μ = filtrate viscosity*, poise
c = feed solids concentration, g/cu cm
b = slope of θ/V versus V curve, sec/cm^6

*assume viscosity = 1.002×10^{-2} poise if capture exceeds 99%

Sample Calculation

P = 686 g/sq cm
A = 38.5 sq cm
$\mu = 1.002 \times 10^{-2}$ poise
c = 53.2 g/ℓ
b = slope = .039 sec/cm^6

$$R = \frac{(2)(686)(38.5)^2}{(.01002)(.0532)}\ (0.39)$$

$R = 1.49 \times 10^8\ sec^2/g$
$R = 1.46 \times 10^{11}\ cm/g$

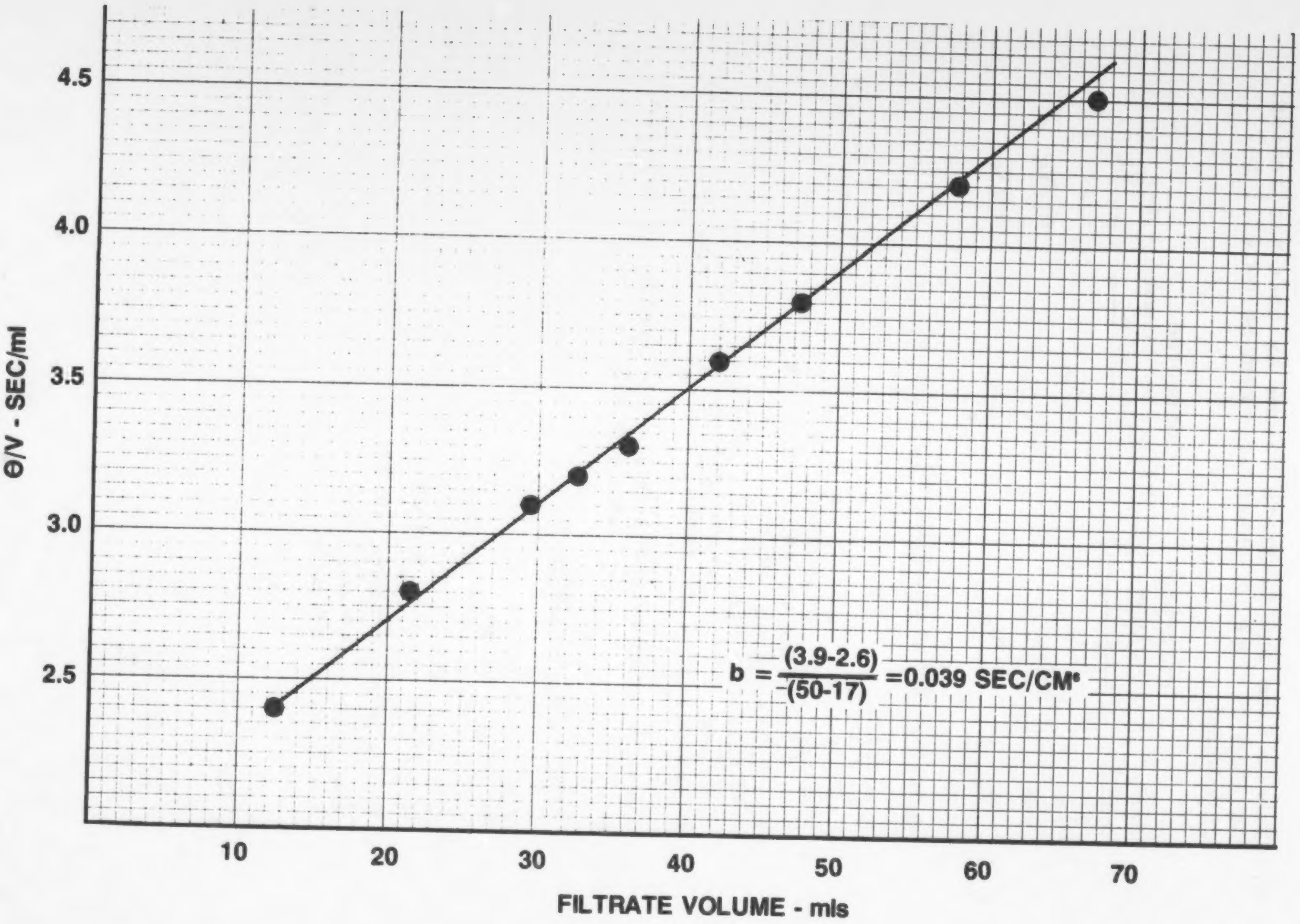

FIGURE B1. SAMPLE PLOT OF SPECIFIC RESISTANCE DATA.

APPENDIX C

LABORATORY ANALYSES AND SCHEDULED TESTS

ANALYTICAL PROCEDURES FOR CHEMICAL ANALYSES

TOTAL PHOSPHORUS:ORTHOPHOSPHATE

Standard Methods, 13th Edition, pp. 526, 532-534. Total phosphorus is determined by the persulfate digestion method. Orthophosphate is determined by the ascorbic acid method. Soluble total phosphorus and soluble orthophosphate are defined as that phosphorus or phosphate material that passes through a 0.45 micron membrane filter.

TOTAL HARDNESS

Standard Methods, 13th Edition, pp. 178-184

METHYL ORANGE ALKALINITY

Standard Methods, 13th Edition, pp. 370-376

BOD

Standard Methods, 13th Edition, pp. 489-495

COD

Standard Methods, 13th Edition, pp. 495-499.

Note: Soluble BOD and soluble COD are defined as material passing through a 0.45 micron membrane filter.

pH

The pH will be determined by using a laboratory pH meter.

TURBIDITY

Turbidity of all samples will be determined with a Hach Chemical Company, Model 2100, laboratory turbidimeter. This device is a nephelometer (90 degree light scattering) which is standardized using standards manufactured by Hach Chemical Company. Turbidity will be reported in NTU.

SUSPENDED SOLIDS

Suspended solids are defined as material retained on a 0.45 micron membrane filter. Concentrated samples of chemical-sewage sludge are filtered through glass mat filter pads (Type GFB). A calibrated 30 mℓ syringe is used to measure sludge sample volumes. The Standard Methods, 13th Edition, pp. 537-538, procedure for Nonfiltrable Residue is used. Volatile and ash content of chemical sludges is determined according to Standard Methods, 13th Edition, p. 536.

IRON

Iron is determined using an atomic absorption analysis on the filtrate generated during the inerts analysis. A known volume of filtrate is diluted using distilled water.

"Techtron Atomic Absorption Manual"

ALUMINUM

Aluminum is determined using an atomic absorption analysis on the filtrate generated during the inerts analysis. A known volume of filtrate is diluted using distilled water containing ionization depressant KCl.

"Techtron Atomic Absorption Manual"

CALCIUM

Calcium is determined using an atomic absorption analysis on the filtrate generated during the inerts analysis. A known volume of filtrate is diluted using CO_2-free water, containing ionization depressant KCl and Lanthanum as a releasing agent.

"Techtron Atomic Absorption Manual"

INERTS

Inerts are determined by digesting the ash with 70 mℓ of 2:1 $HCl:HNO_3$. Following the digestion, filtration through a 0.45 micron filter is performed. All material retained on the filter is reported as inert material.

SCHEDULED ANALYTICAL TESTS OF PROCESS STREAM COMPOSITE SAMPLES

Water Quality Parameter	Raw Wastewater	Chemical Treatment Effluent
Turbidity	a	a
Suspended Solids	a	a
Total and Soluble Phophorus	a	a
Chemical Oxygen Demand	a	a
Soluble Chemical Oxygen Demand	a	a
Biochemical Oxygen Demand	a	a
Soluble Biochemical Oxygen Demand	a	a
pH	a	a
Methyl Orange Alkalinity	a	a
Al	-	b
Fe	-	b

Dewatering Runs	Parameter	
Routine Sludge Inventories	Total Solids	c
Vacuum Filter and Pressure Filter		
Feed	Total Solids	d
Feed and Lime	Total Solids	d
Filtrate	Total Suspended Solids	d
Cake	% Solids, % Volatile, % Ash, % Inert, % Fe or Al, % P	d
Centrifuge		
Feed	Total Solids	d
Centrate	Total Solids	d
Cake	% Solids, % Volatile % Ash, % Inert, % Fe or Al, % P	d

a - Routine analysis on 48 hour composites
b - Rarely
c - As used or discarded
d - On demand

APPENDIX D: SAMPLE JAR TEST DATA SHEET

Test No. ____________________

Company ____________________

Date Tested ____________________

Address ____________________

____

By ____________________

Location ____________________

Material **Solids Consisting of** ____________________

Liquids Consisting of ____________________

Object of Test ____________________

____

RAW SAMPLE

TURBIDITY, JTU ____________ ____________

SUSPENDED SOLIDS, mg/l ____________ ____________

pH, Units ____________ ____________

Color, Units ____________ ____________

Sample Size ____________________

Container Size ____________________

Temp. °C ____________________

Mix Time Flash ____________ Min. @ ________ RPM

Floc. ____________ Min. @ ________ RPM

Settling ____________ Min.

SAMPLE NUMBER		1	2	3	4	5	6	**NOTES**
______	Dosage, mg/l							
______ mg/ml	Addition, ml							
______	Dosage, mg/l							
______ mg/ml	Addition, ml							
______	Dosage, mg/l							
______ mg/ml	Addition, ml							
______	Dosage, mg/l							
______ mg/ml	Addition, ml							
Interface Height (In In. @ × Min.)	Min.							
	Min.							
	Min.							
	Min.							
	Min.							
	Min.							
Height of Sample	In.							
Height After	Min.							
Difference	In.							
Rate	In./Min.							
Rise Rate (gpm/sq ft) = (Rate)(0.312)*								
* Includes 0.50 Scale Up Factor								
SUPERNATANT CHARACTERISTICS		@ ______ **Min.**						
Turbidity, JTU								
Suspended Solids, mg/l								
Color, Units								
pH, Units								
Sludge Generation, mg/l								
	lbs/MG							
Sludge Volume	mls							
	%							

Form E-3022

APPENDIX E

SLUDGE PRODUCTION DETERMINATIONS

PHASE 1A

$$\frac{\text{Kg DS}}{\text{day}} = 0.864\ Q\ \{\text{TSS} + (P)(3.94)\ \text{mg Al PO}_4/\text{mg P}) + (\text{Al}_r)(2.89\ \text{mg Al(OH)}_3/\text{mg Fe})\} \qquad (1)$$

$$Q = 1.40\ \ell\text{ps}$$
$$\text{TSS} = 66\ \text{mg}/\ell$$
$$P = 4.4\ \text{mg}/\ell$$
$$\text{Al dosage} = 14.7\ \text{mg}/\ell$$
$$\text{Al}_r = \text{Al dosage} - (\text{Al used in Al PO}_4)$$

$$\text{Al used in Al PO}_4 = \frac{4.4\ \text{mg P}}{\ell} \times \frac{.87\ \text{mg Al}}{\text{mg P}} = \frac{3.8\ \text{mg Al}}{\ell}$$
$$\text{Al}_r = 14.7 - 3.8 = 10.9\ \text{mg}/\ell$$

$$\frac{\text{Kg DS}}{\text{day}} = (0.864)(1.40)\ \{66 + (4.4)(3.94) + (10.9)(2.89)\}$$
$$= 13.9\ \frac{\text{Kg DS}}{\text{day}}$$

PHASE 1F

$$\frac{\text{Kg DS}}{\text{day}} = .0864\ Q\ \{\text{TSS} + (P)(4.87\ \text{mg FePO}_4) + (\text{Fe}_r)(1.89\ \text{mg Fe(OH)}_3/\text{mg Fe})\} \qquad (2)$$

$$Q = 1.39\ \ell\text{ps}$$
$$\text{TSS} = 66\ \text{mg}/\ell$$
$$P = 4.4\ \text{mg}/\ell$$
$$\text{Fe dosage} = 28.3\ \text{mg}/\ell$$
$$\text{Fe}_r = \text{Fe dosage} = (\text{Fe used in FePO}_4)$$

$$\text{Fe used in FePO}_4 = \frac{4.4\ \text{mg P}}{\ell} \times \frac{1.81\ \text{mg Fe}}{\text{mg P}} = \frac{8.0\ \text{mg Fe}}{\ell}$$
$$\text{Fe}_r = 28.3 - 8.0 = 20.3\ \text{mg}/\ell$$

$$\frac{\text{Kg DS}}{\text{day}} = (.0864)(1.39)\ \{66 + (4.4)(4.87) + (20.3)(1.89)\}$$
$$= 15.0\ \frac{\text{Kg DS}}{\text{day}}$$

PHASE 2A

$$\frac{\text{Kg DS}}{\text{day}} = .0864\ Q\ \{TSS + (P)(3.49) + (Al_r)(2.89)\} \qquad (1)$$

$Q = 2.56\ \ell ps$

$TSS = 61\ mg/\ell$

$P = 5.7\ mg/\ell$

$Al\ dosage = 15.9\ mg/\ell$

$Al_r = Al\ dosage - (Al\ used\ in\ AlPO_4)$

$$Al\ used\ in\ AlPO_4 = \frac{5.7\ mg\ P}{\ell} \times \frac{.87\ mg\ Al}{mg\ P} = \frac{5.0\ mg\ Al}{\ell}$$

$$Al_r = 15.8 - 5.0 = \frac{10.9\ mg\ Al}{\ell}$$

$$\frac{\text{Kg DS}}{\text{day}} = (.0864)(2.56)\ \{61 + (5.7)(3.94) + (10.9)(2.89)\}$$

$$= 25.4\ \frac{\text{Kg DS}}{\text{day}}$$

PHASE 2F

$$\frac{\text{Kg DS}}{\text{day}} = .0864\ Q\ \{TSS + (P)(4.87) + (Fe_r)(1.89)\} \qquad (2)$$

$Q = 2.56\ \ell ps$

$TSS = 61\ mg/\ell$

$P = 5.7\ mg/\ell$

$Fe\ dosage = 32.8\ mg/\ell$

$Fe_r = Fe\ dosage = (Fe\ used\ in\ FePO\ )$

$$Fe\ used\ in\ FePO_4 = \frac{5.7\ mg\ P}{\ell} \times \frac{1.81\ mg\ Fe}{mg\ P} = \frac{10.3\ mg\ Fe}{\ell}$$

$$Fe_r = 32.8 = 10.3 = 22.5\ mg\ Fe/\ell$$

$$\frac{\text{Kg DS}}{\text{day}} = (.0864)(2.56)\ \{61 + (5.7)(4.87) + (2..5)(1.89)\}$$

$$= 29.2\ \frac{\text{Kg DS}}{\text{day}}$$

PHASE 3A

$$\frac{\text{Kg DS}}{\text{day}} = .0864\ Q\ \{TSS + (P)(3.94) + (Al_r)(2.89)\} \qquad (1)$$

$Q = 2.47\ \ell ps$

$TSS = 108\ mg/\ell$

$P = 5.4\ mg/\ell$

$Al\ dosage = 8.1\ mg/\ell$

$Al_r = Al\ dosage - (Al\ used\ in\ AlPO_4)$

$$Al\ used\ in\ AlPO_4 = \frac{5.4\ mg\ P}{\ell} \times \frac{.87\ mg\ Al}{mg\ P} = \frac{4.7\ mg\ Al}{\ell}$$

$$Al_r = 8.1 - 4.7 = \frac{3.4\ mg\ Al}{\ell}$$

$$\frac{\text{Kg DS}}{\text{day}} = (.0864)(2.47)\ \{108 + (5.4)(3.94) + (3.4)(2.89)\}$$
$$= 29.7\ \frac{\text{Kg DS}}{\text{day}}$$

Phase 3F-I

$$\frac{\text{Kg DS}}{\text{day}} = .0864\ Q\ \{\text{TSS} + P\ (4.87) + (\text{Fe}_r)(1.89)\} \qquad (2)$$

$$Q = 2.44\ \ell\text{ps}$$
$$\text{TSS} = 88\ \text{mg}/\ell$$
$$P = 5.2\ \text{mg}/\ell$$
$$\text{Fe dosage} = 22.8$$
$$\text{Fe}_r = \text{Fe dosage} - (\text{Fe used in FePO}_4)$$
$$\text{Fe used in FePO}_4 = \frac{5.2\ \text{mg P}}{\ell} \times \frac{1.81\ \text{mg Fe}}{\text{mg P}} = \frac{9.4\ \text{mg Fe}}{\ell}$$
$$\text{Fe}_r = 22.8 - 9.4 = 13.4\ \text{mg Fe}$$

$$\frac{\text{Kg DS}}{\text{day}} = (.0864)(2.44)\ \{88 + (5.2)(4.87) + (13.4)(1.89)\}$$
$$= 29.1\ \frac{\text{Kg DS}}{\text{day}}$$

PHASE 3F-II

$$\frac{\text{Kg DS}}{\text{day}} = .0864\ Q\ \{\text{TSS} + P\ (4.87) + (\text{Fe}_r)(1.89)\} \qquad (2)$$

$$Q = 2.47\ \ell\text{ps}$$
$$\text{TSS} = 115\ \text{mg}/\ell$$
$$P = 5.5\ \text{mg}/\ell$$
$$\text{Fe dosage} = 10.9\ \text{mg}/\ell$$
$$\text{Fe}_r = \text{Fe dosage} - (\text{Fe used in FePO}_4)$$
$$\text{Fe used in FePO}_4 = \frac{5.5\ \text{mg P}}{\ell} \times \frac{1.81\ \text{mg Fe}}{\text{mg P}} = \frac{10.0\ \text{mg Fe}}{\ell}$$
$$\text{Fe}_r = 10.9 = 10.0 = 0.9\ \text{mg Fe}/\ell$$

$$\frac{\text{Kg DS}}{\text{day}} = (0.0864)(2.47)\ \{115 + 5.5\ (4.87) + (0.9)(1.89)\}$$
$$= 30.7\ \frac{\text{Kg DS}}{\text{day}}$$

PHASE 4A

$$\frac{\text{Kg DS}}{\text{day}} = (.0864)\ Q\ \{\text{TSS} + (P)(3.94) + (\text{Al}_r)(2.89)\} \qquad (1)$$

$$Q = 2.49\ \ell\text{ps}$$
$$\text{TSS} = 120\ \text{mg}/\ell$$
$$P = 5.8\ \text{mg}/\ell$$
$$\text{Al dosage} = 15.6\ \text{mg}/\ell$$
$$\text{Al}_r = \text{Al dosage} - (\text{Al used in AlPO}_4)$$
$$\text{Al used in AlPO}_4 = \frac{5.8\ \text{mg P}}{\ell} \times \frac{.87\ \text{mg Al}}{\text{mg P}} = \frac{5.0\ \text{mg Al}}{\ell}$$
$$\text{Al}_r = 15.6 - 5.0 = \frac{10.6\ \text{mg Al}}{\ell}$$

$$\frac{\text{Kg DS}}{\text{day}} = (.0864)(2.49)\{120 + (5.8)(3.94) + (10.6)(2.89)\}$$
$$= 37.4\ \frac{\text{Kg DS}}{\text{day}}$$

APPENDIX F

SLUDGE CHARACTERIZATION DETERMINATIONS

Sample Calculation

PHASE 3F

TSS = 108 mg/ℓ
Voltailes = 82%
P = 5.4 mg/ℓ
Fe dosage = 13.3 mg/ℓ

Constituents Present

Volatiles = (108 mg/ℓ)(.82) = 89 mg/ℓ VSS
Inerts = (108 mg/ℓ)(.18) = 19 mg/ℓ Inerts

$$\mathrm{FePO_4} = \left(\frac{5.4\ \mathrm{mg\ P}}{\ell}\right)\left(\frac{4.87\ \mathrm{mg\ FePO_4}}{\mathrm{mg\ P}}\right) = 26\ \mathrm{mg}/\ell\ \mathrm{FePO_4}$$

$$\mathrm{Fe(OH)_3} = \left(\frac{3.5\ \mathrm{mg\ Fe}}{\ell}\right)\left(\frac{1.89\ \mathrm{mg\ Fe(OH)_3}}{\mathrm{mg\ Fe}}\right) = 7\ \mathrm{mg}/\ell\ \mathrm{Fe(OH)_3}$$

TOTAL = 141 mg/ℓ

Constituents Present by Percent

Volatiles = 89/141 = 63
Inerts = 19/141 = 14
$\mathrm{FePO_4}$ = 26/141 = 18
$\mathrm{Fe(OH)_3}$ = 7/141 = 5 23% chemical
100

APPENDIX G

COST ANALYSIS ASSUMPTIONS

This Appendix contains a detailed summary of the assumptions made in performing the cost analysis discussed in Section XII.

I. WASTEWATER

  A. Flows

    1. 20,000 cu m/day
    2. 40,000 cu m/day
    3. 200,000 cu m/day

  B. Strength

    1. BOD_5 = 200 mg/ℓ
    2. TSS = 200 mg/ℓ
    3. VSS = 160 mg/ℓ
    4. Phosphorus = 8 mg P/ℓ

II. CHEMICAL TREATMENT

  A. Primary coagulants

    1. Alum (8.25% as Al_2O_3)
    2. Ferric chloride (43.4% as $FeCl_3$)

  B. Phosphorus removal levels

    1. 80 percent phosphorus removal
    2. 95 percent phosphorus removal

  C. Primary coagulant doses

    1. Alum

      a. 12 mg Al/ℓ at 80 percent phosphorus removal
      b. 22 mg Al/ℓ at 95 percent phosphorus removal

    2. Ferric chloride

      a. 22 mg Fe/ℓ at 80 percent phosphorus removal
      b. 40 mg Fe/ℓ at 95 percent phosphorus removal

D. Sludges produced (by calculation)

1. Alum-primary
   a. 80 percent phosphorus removal
      1. 0.221 Kg DS/cu m
      2. 69.1 percent volatile solids
      3. 12.2 percent inert solids
      4. 18.7 percent chemical solids

   b. 95 percent phosphorus removal
      1. 0.247 Kg DS/cu m
      2. 61.8 percent volatile solids
      3. 10.9 percent inert solids
      4. 27.3 percent chemical solids

2. Ferric-primary sludge
   a. 80 percent phosphorus removal
      1. 0.228 Kg DS/cu m
      2. 67.2 percent volatile solids
      3. 11.8 percent inert solids
      4. 21.0 percent chemical solids

   b. 95 percent phosphorus removal
      1. .258 Kg DS/cu m
      2. 59.2 percent volatile solids
      3. 10.4 percent inert solids
      4. 30.4 percent chemical solids

III. SLUDGE THICKENING

A. Alum-primary sludge

1. Underflow = 4.6 percent total solids at 80 percent phosphorus removal

2. Underflow = 3.8 percent total solids at 95 percent phosphorus removal

B. Ferric-primary sludge

1. Underflow = 5.6 percent total solids at 80 percent phosphorus removal

2. Underflow = 5.4 percent total solids at 95 percent phosphorus removal

IV. SLUDGE DEWATERING

A. Vacuum filtration

1. Alum-primary sludge
   - a. 80 percent phosphorus removal
      1. Cycle time = 5.0 minutes
      2. Lime conditioning dose = 25 percent
      3. Filtration rate = 13.5 Kg/hr-sq m (exc. chemicals)
      4. Cake = 25.5 percent dry solids
      5. Cake = 55 percent volatiles
   - b. 95 percent phosphorus removal
      1. Cycle time = 5.0 minutes
      2. Lime conditioning dose = 25 percent
      3. Filtration rate = 8.6 Kg/hr-sq m (exc. chemicals)
      4. Cake = 26.5 percent dry solids
      5. Cake = 49 percent volatiles
2. Ferric-primary sludge
   - a. 80 percent phosphorus removal
      1. Cycle time = 3.3 minutes
      2. Lime conditioning dose = 30 percent
      3. Filtration rate = 10.2 Kg/hr-sq m (exc. chemicals)
      4. Cake = 34.5 percent dry solids
      5. Cake = 52 percent volatiles
   - b. 95 percent phosphorus removal
      1. Cycle time = 3.3 minutes
      2. Lime conditioning dose = 30 percent
      3. Filtration rate = 8.7 Kg/hr-sq m (exc. chemicals)
      4. Cake = 34.5 percent dry solids
      5. Cake = 46 percent volatiles

B. Centrifugation

1. Alum-primary sludge
   - a. 80 percent phosphorus removal
      1. Total solids capture = 95 percent
      2. Flow at P600 = 8.7 ℓpm
      3. Polymer dose = 0.30 percent
      4. Cake = 18 percent dry solids
      5. Cake = 69 percent volatiles

b. 95 percent phosphorus removal

1. Total solids capture = 95 percent
2. Flow at P600 = 8.7 ℓpm
3. Polymer dose = 0.30 percent
4. Cake = 18 percent dry solids
5. Cake = 62 percent volatiles

2. Ferric-primary sludge

a. 80 percent phosphorus removal

1. Total solids capture = 90 percent
2. Flow at P600 = 6.7 ℓpm
3. Polymer dose = 0.35 percent
4. Cake = 22 percent dry solids
5. Cake = 69 percent volatiles

b. 95 percent phosphorus removal

1. Total solids capture = 90 percent
2. Flow at P600 = 6.7 ℓpm
3. Polymer dose = 0.35 percent
4. Cake = 22 percent dry solids
5. Cake = 62 percent volatiles

C. Operating schedule (nominal)

1. Vacuum filtration

a. 15 minutes start-up
b. 30 minute shut-down
c. 8 hrs/day at 20,000 cu m/day
d. 16 hrs/day at 40,000 cu m/day
e. 16 hrs/day at 200,000 cu m/day

2. Centrifugation

a. 15 minute start-up
b. 15 minutes shut-down
c. 8 hrs/day at 20,000 cu m/day
d. 16 hrs/day at 40,000 cu m/day
e. 16 hrs/day at 200,000 cu m/day

V. SLUDGE DISPOSAL

A. Multiple hearth incineration

1. Fuel

a. Fuel selected = fuel oil
b. Fuel value = $2.80/10^6 BTU

2. Time of operation

For all cases, it was assumed the furnace would operate for the same time period as the respective unit(s) of dewatering equipment.

3. Design basis

Envirotech technical data

VI. AMORTIZATION, INTEREST, POWER

A. Amortization

20 year period

B. Interest rate

8.0 percent

C. Electricity costs

a. \$.0275/kwh at 20,000 m^3/day
b. \$.025/kwh at 40,000 m^3/day
c. \$.020/kwh at 200,000 m^3/day